PB-222 251

TERTIARY TREATMENT OF COMBINED DOMESTIC AND INDUSTRIAL WASTES

John W. Lee, Jr.

Cornell, Howland, Hayes and Merryfield

Prepared for:

Environmental Protection Agency

May 1973

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Tertiary Treatment of Combined Domestic and Industrial Wastes

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SELECTED WATER RESOURCES ABSTRACTS
INPUT TRANSACTION FORM

1. Report No.
2.
3. Accession No. W 73-12512

4. Title
Tertiary Treatment of Combined Domestic and Industrial Wastes, [illegible]

5. Report Date
6.
8. Performing Organization Report No.

7. Author(s)
John W. Lee, Jr.

10. Project No. EPA Project 11060 DLF

9. Organization
Cornell, Howland, Hayes and Merryfield, Inc.,
~~1600 S.W. Western Boulevard~~
Corvallis, Oregon 97[illegible]

11. Contract/Grant No.
~~[illegible]~~

13. Type of Report and Period Covered

~~12. Sponsoring Organization~~
~~City of Tualatin, Oregon 97062~~

15. Supplementary Notes
Environmental Protection Agency Technology Series report
~~number~~, EPA-R2-73-236, May 1973. 203p, 67 fig, 14 tab, 38 ref

16. Abstract

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The system was designed for an average daily flow of 280,000 gpd and a BOD_5 load of 630 pounds per day. The extended aeration activated sludge process with a design detention time of 24 hours was employed for secondary treatment. An experimental 60 degree inclined tube settler located in the aeration-surge basin provided secondary effluent clarification.

The tertiary system consisted of a four step process: 1) alum and polyelectrolyte coagulation, 2) flocculation, 3) inclined tube sedimentation, and 4) mixed media filtration. The tertiary system demonstrated the capability to produce an effluent quality of less than 10 mg/l BOD_5 and 5 mg/l suspended solids with a total phosphate residual of 0.1 to 1.0 mg/l (as P).

The total capital cost of the facility was $245,800. Based on total annual cost, the cost of treatment at the design conditions was $0.42 per 1000 gallons processed and $0.19 per pound of BOD_5 removed. (EPA)

~~This report was submitted in fulfillment of Grant No. 11060 DLF, under the partial sponsorship of the Environmental Protection Agency.~~

17a. Descriptors
*Waste water Treatment, *Aerobic Treatment, *Tertiary Treatment, *Industrial Wastes, Activated Sludge, *Domestic Wastes, Flocculation, Filtration, Chemical Precipitation, Phosphates, Treatment facilities, *Oregon
Biochemical oxygen demand

17b. Identifiers
Pet Food Processing Wastes, Mixed Media Filter, Tube Clarifier, Alum, BOD5, Polyelectrolyte, Treatment Costs, Phosphate Removal, Combined treatment,
*Tualatin (Ore)

Copy available from GPO Sup Doc as EP1.23/2: 73-236, $2.60 ; microfiche from NTIS as PB 222 251, $0.95.

17c. COWRR Field & Group 05D

18. Availability
19. Security Class. (Report)
20. Security Class. (Page)
21. No. of Pages
22. Price

Send To:
WATER RESOURCES SCIENTIFIC INFORMATION CENTER
U.S. DEPARTMENT OF THE INTERIOR
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Corvallis, Oregon

WRSIC 102 (REV. JUNE 1971)

1a

EPA-R2-73-236
May 1973

TERTIARY TREATMENT
OF
COMBINED DOMESTIC AND INDUSTRIAL WASTES

by

John W. Lee, Jr.

Grant No. 11060 DLF
Program Element 1B2037
(Project 12130 DLF)

For sale by the Superintendent of Documents, U.S. Government Printing Office, Washington, D.C. 20402
Price $2.00 domestic postpaid or $1.25 GPO Bookstore

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Prepared for
OFFICE OF RESEARCH AND MONITORING
U.S. ENVIRONMENTAL PROTECTION AGENCY
WASHINGTON, D. C. 20460

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ABSTRACT

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SECTION I
CONCLUSIONS

The operation of a small (0.28 mgd) secondary-tertiary facility, providing treatment of combined domestic and pet food manufacturing industrial wastewaters for the City of Tualatin, Oregon, was evaluated from August 1970 through October 1971.

The following conclusions have been reached, based on the results of the study presented in this report:

1. Application of alum coagulation, sedimentation, and mixed-media filtration is an effective tertiary treatment process for polishing and removing phosphate from the effluent of a complete-mix activated sludge extended aeration system.

2. A final effluent quality of 10 mg/l BOD_5, 5 mg/l total suspended solids, and 0.1 to 1.0 total phosphate could be maintained, even under high suspended solids conditions in the secondary effluent.

3. High influent organic loadings exceeded the design aeration capacity for much of the demonstration period, substantially affecting the performance of the secondary treatment system.

4. The tertiary effluent quality reported above was maintained while treating secondary effluent from an extended aeration system having an average sludge age of 13.3 days (range 4 to 31 days), and an average aeration-surge basin detention time of 2.6 days (range 1.6 to 7.1 days).

5. The influent suspended solids had a significant effect upon the gross sludge yield from the secondary treatment system. The influent suspended solids did not appear to begin biological breakdown until retained in the secondary system for at least five days. Eighty (80) percent of the influent suspended solids was determined to be resistant to biological degradation.

6. The gross sludge yield, nondegraded influent suspended solids plus biologically synthesized solids in the MLSS, of the secondary system at a sludge age of 10 to 30 days was determined to be about 1.0 pounds of solids per pound of influent BOD_5 removed.

7. The net sludge yield, solids biologically synthesized less endogenous respiration loss, decreased from 0.4 to 0.17 pounds per pound of BOD_5 removed between sludge ages of 5 and 30 days.

8. The net sludge yield and endogenous coefficients were determined to be 0.60 and 0.15 day^{-1} respectively.

9. The substrate removal coefficient, k, based on total MLSS was determined to be 0.0235 lbs BOD_5 removed per day per lbs of MLSS per mg/l soluble effluent BOD_5.

10. The substrate removal coefficient, k' , expressed in terms of the estimated biological mass (M_B), was calculated to be 0.098 lbs BOD_5 removed per day per lbs M_B per mg/l effluent soluble BOD_5.

11. The secondary tube clarifier performance was found to be dependent upon MLSS concentration, overflow rate, temperature, pH, and the extent of nitrification.

12. Complete nitrification (98 percent oxidation of soluble nitrogen) occurred at D.O. levels above 1.0 to 1.5 mg/l and sludge ages greater than 10 to 12 days within a temperature range of 8 to 22 degrees C.

13. Sludge ages of less than 5 days or D.O. levels of less than 0.4 to 0.6 mg/l effectively inhibited nitrification.

14. Optimum phosphate removal (approaching 60 percent) in the aeration-surge basin occurred at a pH of 7.0 to 7.25.

15. The aerator field oxygen transfer rate was determined to be about 1.4 lbs O_2/hp-hr. The corresponding clean water transfer rate was calculated to be about 2.1 lbs of O_2/hp/hr.

16. The filter throughput volume per cycle was not affected by filtration rates ranging from 1.4 to 4.0 gpm/sq ft.

17. Solids carryover from the tertiary tube settler significantly reduced the tertiary filter cycle time.

18. Filter cycle time was also influenced by the secondary effluent suspended solids concentration, alum dosage, and excessive prefilter polyelectrolyte addition.

19. Anionic polyelectrolyte addition of up to 2.0 mg/l prior to flocculation did not appear to appreciably aid chemical floc formation or improve tertiary performance.

20. Anionic polyelectrolyte dosages of 0.03 to 0.06 mg/l applied prior to filtration, were effective in controlling turbidity breakthrough. Prefilter polymer dosages in excess of 0.1 mg/l produced "filter binding," reducing the filter cycle time.

21. Orthophosphate removal in the tertiary system was dependent upon the alum dosage, pH, and the secondary effluent orthophosphate concentration.

22. Optimum phosphate removals occurred at a tertiary effluent pH of 5.5 to 6 and at [Al(III)]/[secondary effluent ortho P] molar concentration ratios of 2 to 3.

23. The lowest levels of residual suspended solids and turbidity in the tertiary effluent occurred in a pH range of 5.8 to 6.3. Tertiary effluent turbidity also appeared to correlate with nitrate concentration.

24. The extent of nitrification and its effect on alkalinity in the aeration-surge basin had a significant effect on tertiary operation.

25. Lagooning of waste activated and tertiary chemical sludge was an effective means of storing waste solids. No objectionable odors were detected from the lagoons. The volatile solids content of the waste activated and chemical sludges after storage for 6 to 9 months averaged 52 percent and 48.5 percent, respectively.

26. Separate drying bed experiments indicated the stabilized waste activated sludge applied at an average depth of 6 inches, would dry to a cinder-like humus material with a moisture content of 9 to 12 percent after 4 weeks.

27. Stabilized chemical sludge applied to a small-scale drying bed at a depth of 9 inches and 3.3 percent solids content dewatered to a solids content of 68 percent after 27 days.

28. A combined secondary-tertiary treatment system of this type can be constructed and operated at a substantial savings in cost compared to a conventional activated sludge system with tertiary treatment. The total capital cost was about 71 percent of the cost for a comparable size conventional activated sludge plant with tertiary treatment.

29. The total annual cost, based on design conditions, was calculated to be \$0.421 per 1,000 gallons treated. The individual cost allocations for the secondary and tertiary systems were estimated to be \$0.287 and \$0.134 per 1,000 gallons treated respectively. (These costs do not include waste solids disposal).

30. In terms of organic loading, the total annual treatment cost based on design conditions averaged \$0.187 per pound of BOD_5 removed for the first year of operation. (This cost does not include waste solids disposal).

SECTION II
RECOMMENDATIONS

The demonstration program identified several design modifications and items of additional equipment, which can improve plant performance and operation. Additional operational research studies to further explore various aspects of the secondary and tertiary treatment processes would also be desirable.

DESIGN MODIFICATIONS

1. Gas accumulation under the aeration-surge basin PVC liner caused the membrane to tear away from the ringwall footing and float. The liner was repaired and a venting system installed under the liner to provide gas relief. It is recommended that a gas venting system be included in the design of aeration-surge basins using membrane liners.

2. Where wide fluctuations in organic loading are expected in extended aeration systems treating combined industrial and domestic wastewater, it is recommended that two speed, as opposed to single speed, mechanical surface aerators be considered. In addition it is suggested that more units of lower horsepower be used than a lesser number of higher aeration capacity. This concept allows flexibility in adjusting the air supply to satisfy the incoming oxygen demand. Increased equipment costs can be offset by the savings in power realized.

3. Satisfactory secondary tube clarifier performance was restricted to MLSS levels below 2,200 mg/l. Additional study of tube module design to extend the allowable MLSS range of effective clarification is warranted. Provisions were not included to thicken waste activated sludge. A means of concentrating waste activated sludge should be incorporated in the secondary tube clarifier design. An air sparging system, installed midway through the demonstration project, proved to be an effective method of cleaning and preventing solids buildup in the tubes. It is recommended that air sparging also be included in the clarifier design.

4. Considerable short circuiting occurred in the flocculator tank. Modifications were made to minimize this condition. It is recommended that in future designs of the flocculation system, careful consideration be given to preventing short circuiting. For small tertiary systems it may be convenient to include the flocculator as part of the tertiary tube settler/filter unit.

5. High velocity constrictions in the tertiary tube settler inlet system were suspected to cause a breakup of the fragile chemical floc. The results of modifications to correct this condition were inconclusive. It is recommended that the inlet system be redesigned to prevent turbulence from damaging the chemical floc.

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6. The tertiary tube settler modules did not have adequate storage volume for the solids loadings encountered. The effective surface overflow rate should be reduced to about 35 gallons per day per square foot.

7. Pumping the chemical sludge holding tank supernatant, after settling, to the aeration-surge basin resulted in severe pH depression when the secondary system was highly nitrified. It is recommended that the supernatant be combined with secondary effluent for return to the tertiary system.

8. Low secondary effluent alkalinity conditions, produced by nitrification in the aeration-surge basin, required the addition of soda ash to the tertiary influent to maintain pH control in the tertiary process. It is recommended that soda ash mixing, storage and metering equipment be included as part of the chemical feed facilities in secondary-tertiary treatment systems of this type treating relatively soft waters.

ADDITIONAL EQUIPMENT

1. The plant effluent was the only wastewater flow accurately measured and recorded. Because of low influent flow conditions and the surge capacity of the aeration-surge basin, the plant influent and effluent flow rates were not necessarily equivalent. Measurement and recording of the plant influent, as well as waste activated sludge and sludge storage lagoon supernatant return flows, are recommended.

2. For treatment systems receiving combined domestic and industrial wastewaters with large fluctuations in both organic and hydraulic loadings, flow proportioned automatic sampling of the plant influent, secondary effluent and tertiary effluent is recommended. Flow proportioned automatic sampling of significant industrial waste flows is also suggested.

3. Tertiary filter head loss was indicated, but not recorded. A recording turbidimeter was borrowed during the demonstration program to monitor tertiary effluent quality. It is recommended that both filter head loss and tertiary effluent turbidity be recorded, as well as indicated, to effectively regulate chemical feed, assess tertiary system performance and control plant effluent quality.

4. The most important control parameter for the tertiary process was pH. It is recommended that the pH of the flocculator effluent be continuously indicated and recorded.

5. The extent of nitrification in the secondary system substantially influenced tertiary treatment performance. Where size of the treatment facility can justify the expense, it is suggested that the D.O. and ammonium ion concentration of the aeration-surge basin be measured and recorded.

AREAS FOR FURTHER STUDY

1. The effect of polymer addition on flocculation was not adequately demonstrated and should be further investigated.

2. Alkalinity is suspected to influence alum-phosphate coagulation and warrants further research.

3. Tertiary effluent quality depended primarily on the characteristics of the secondary effluent and chemical dosages applied, and did not appear to be affected by filtration rate. Additional study is needed to verify these observations, where the tertiary influent characteristics can be more closely controlled.

5. Ammonium ion, pH and the oxidative condition in the aeration-surge basin appeared to affect secondary clarifier performance. Further research is needed to more fully explore the effect of these variables on coagulation of colloidal particulate.

6. Optimum orthophosphate removals in the secondary system were observed to occur in a narrow pH range. Further research is needed to determine the influence of pH and the mechanism(s) of phosphate removal in activated sludge systems.

7. Nitrate ion and pH appeared to affect the turbidity level and suspended solids concentration of the tertiary effluent. This is another area deserving additional study.

SECTION III
INTRODUCTION

SCOPE

A secondary-tertiary system, designed to treat combined domestic and industrial wastes, was constructed at the City of Tualatin, Oregon, and studied during the period from August 1970 through October 1971. This project was financed with the aid of a research, development, and demonstration grant provided by the Environmental Protection Agency (EPA), under grant project No. 11060 DLF.

The basic purpose of this program was to develop design criteria, evaluate operating characteristics, and determine the effectiveness of a small, low cost, municipal tertiary treatment plant treating combined domestic and industrial wastes. At the time this project was initiated, no plant scale installations of this type of secondary-tertiary treatment, providing nutrient removal, existed.

The primary objectives of this program were to:

1. Demonstrate in full scale plant operation, the BOD, suspended solids, and the phosphate removal capabilities of a completely automated high-rate tertiary filtration system treating combined domestic and industrial waste flows.

2. Determine the economics of the system providing an effluent quality of 10 mg/l total BOD and 10 mg/l suspended solids or less and a total phosphate residual of 0.1 to 1.0 mg/l (as P).

3. Demonstrate the applicability of the tertiary treatment process as supplementary treatment for the extended aeration modification of the activated sludge process. Extended aeration is becoming prevalent throughout the country as a secondary treatment process for small municipalities.

4. Demonstrate the application of tube type clarifiers for use in extended aeration activated sludge systems as a substitute for conventional clarifier tanks and mechanisms.

5. Study the operating techniques required and the problems associated with disposal of tertiary treatment chemical sludge, including the degree of sludge dewatering and compaction in storage ponds, and the most desirable means of off-site disposal.

6. Provide detailed design criteria and operations recommendations for use in adapting the secondary-tertiary treatment system to other locations.

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Secondary objectives of the project were to:

1. Determine an aeration-surge basin velocity profile.
2. Determine the dewatering and drying characteristics of the waste activated sludge.
3. Determine the solids accumulation in the bottom of the aeration-surge basin at the end of the program.
4. Determine the volume and density of both the chemical and activated sludge deposits in the holding ponds at the end of the program.
5. Identify potential problems associated with chemical and activated sludge disposal.

BACKGROUND

National awareness of the deteriorating quality of many of our receiving waters has prompted governmental agencies to place increasingly stringent effluent quality discharge requirements on wastewater treatment plants. Secondary treatment alone is rapidly becoming inadequate in many locations of the country and must be supplemented by tertiary treatment.

In addition to the domestic load, many small communities are faced with the problem of providing wastewater treatment services to local industry. The waste loads from such industries as food processors can constitute the major organic load on the treatment plant, adding to the difficulty of providing the degree of treatment required to comply with governmental discharge standards. These small cities therefore require an economical treatment system that will provide the level of treatment necessary, without placing a financial burden on the community.

Severe pollution problems encountered during periods of low stream flow prompted the Oregon State Sanitary Authority (OSSA), presently the Oregon State Department of Environmental Quality (OSDEQ), to issue a directive in September 1966 prohibiting construction of additional secondary treatment plants along the Tualatin River, unless discharge to the river was eliminated during the summer months or tertiary treatment provided. Tertiary treatment was defined as an effluent containing not more than 10 mg/l BOD and 10 mg/l suspended solids. As a result of this action by the OSSA, all new construction activity was halted in the Tualatin River Basin.

The City of Tualatin, Oregon, having convenient highway and rail access as well as available land, had actively encouraged industrial and residential development. Since adequate sewerage service was a major factor in attracting new growth, a method of sewage treatment approved by the OSSA was necessary for development to continue. No installations of small municipal tertiary plants existed at that time. A critical need existed for the development of plant design criteria and operating characteristics, demonstration of advanced waste treatment equipment, and economic evaluation of such installations.

In December 1966, the City of Tualatin submitted an application to the Federal Government for a Municipal Research and Demonstration Grant to assist in financing the design, construction, and operation of a demonstration tertiary treatment plant for combined domestic and industrial wastes.

Tentative approval of the grant project was given in November 1968, by the then Federal Water Pollution Control Administration (FWPCA), presently the EPA. The original grant project No. WPRD 27-01-68 was changed by the FWPCA to No. 11060 DLF.

The City of Tualatin retained the engineering firm of Cornell, Howland, Hayes & Merryfield to design the treatment facility and to exercise technical supervision over the research and development program. Construction was completed, and the facility started up in April 1970. Several problems delayed the beginning of the research and demonstration program until August 1970.

The treatment system received a combination of domestic and commercial wastewaters as well as an industrial effluent from a pet food manufacturer.

The secondary-tertiary treatment system was designed to be able to produce an effluent containing not more than 10 mg/l BOD and 10 mg/l suspended solids and a total phosphate content of 0.1 to 1.0 mg/l (as P). The secondary treatment process used was a completely mixed extended aeration activated sludge system without primary treatment. A tube settler located in the aeration-surge basin provided secondary clarification. The tertiary treatment system was a chemical (alum-polyelectrolyte) coagulation, sedimentation, and filtration process developed by Neptune MicroFLOC, Inc.

The treatment plant was designed and constructed as a temporary facility for research and development in conjunction with treatment of combined industrial and domestic wastes. The plant was to operate for approximately 5 years, at which time connection would be made to a regional sewerage system. Research and demonstration activities were carried out during the first 15 months of operation.

A complete list of definitions of the technical terms used in this report may be found in the WPCF Glossary [1], and a list of abbreviations and symbols used is contained in Section XIII.

THEORETICAL CONSIDERATIONS

EXTENDED AERATION ACTIVATED SLUDGE PROCESS

GENERAL–The complete mix extended aeration activated sludge process differs from mo[illegible]onventional activated sludge systems in several respects. Primary treatment is not usually provided. Incoming raw waste is instantaneously and completely mixed with the aeration basin contents. Larger aeration basins are used to increase the aeration detention time to between 12 and 24 hours or longer.

The effect of the larger aeration volume is to provide a greater total mass of organisms (M) and a longer activated sludge retention time than conventional systems at the same microorganism (mixed liquor) concentrations to metabolize the incoming organic wastes or "food" (F). The longer sludge retention time and lower ratio of F to M result in an overall increase in waste metabolism, reducing the amount of biological solids that must be disposed of.

The larger mass of active microorganisms, as well as the concept of complete mixing, provides inherently greater process stability than conventional activated sludge systems, minimizing the potential for sudden variations in organic and hydraulic loading or introduction of toxic materials to upset the biological treatment process.

A brief discussion of the theory of the basic activated sludge process follows:

MICROBIOLOGY–The living organisms found in activated sludge are classified as either plants or animals. The plants consist of bacteria and fungi. The animals are primarily protozoa, rotifers, and nematodes.

Hawkes [2] states that bacteria are normally the dominant primary feeders on organic wastes. Various holozoic protozoa are secondary feeders. Rotifers and nematodes occupy higher levels in the food chain. Fungi normally cannot compete with bacteria, but they may predominate as primary feeders if certain conditions exist, such as low pH, nitrogen deficiency, or low dissolved oxygen [3]. High-carbohydrate wastes are also reported to stimulate fungi growth.

The composition of the organic wastes determines to a large extent which bacterial genera will predominate [2,3]. Protein wastes favor *Alcaligenes, Flavobacterium, and Bacillus,* while carbohydrate wastes favor *Pseudomonas.* A large population of free swimming bacteria will sustain free swimming ciliata as the predominate protozoa; however, if the food level is lowered by a reduction in the free swimming bacterial population, the free swimming ciliates will yield to stalked ciliates which require less energy.

Rotifers thrive in very stable systems and are better indicators of stable conditions than are the nematode worms.

METABOLISM–The metabolic reactions that occur within activated sludge can be divided into three phases: (1) oxidation, (2) synthesis, and (3) endogenous respiration. The reactions in each phase are described by the following general equations, formulated by Weston and Eckenfelder [4]:

(1) **Organic Matter Oxidation**

$$C_xH_yO_z + aO_2 \rightarrow xCO_2 + bH_2O + \text{Energy}$$

where: $a = x + \frac{y}{\text{[illegible]}}$ [illegible]

$b = \frac{y}{2}$

(2) **Cell Material Synthesis**

$$C_xH_yO_z + NH_3 + dO_2 + \text{Energy} \rightarrow C_5H_7NO_2 + eCO_2 + fH_2O$$

where: $d = x + \frac{y}{4} - \frac{z}{2} - 5$

$e = x - 5$

$f = \frac{y}{2} - 2$

(3) **Cell Material Endogenous Respiration**

$$C_5H_7NO_2 + 5O_2 \rightarrow 5CO_2 + 2H_2O + NH_3 + \text{Energy}$$

Microorganisms oxidize about one-third of the organic matter removed from wastewater directly to carbon dioxide and water. This oxidation process provides the energy necessary to convert the remaining two-thirds of the organic material removed from solution to cell tissue [5]. A portion of the synthesized cell tissue is subsequently oxidized to carbon dioxide, water and other metabolic end products by endogenous respiration.

SUBSTRATE REMOVAL KINETICS–Several authors [6,7,8,9] have developed equations to describe the BOD, or substrate, removal kinetics of the complete-mix activated sludge process. The Michaelis-Menton relationship is the fundamental principle underlying most of these equations and is the basis on which substrate removal kinetics have been evaluated in this study.

The rate of BOD removal, mass of organisms present and soluble effluent BOD concentration in a completely mixed system are related by the Michaelis-Menton equation as follows:

$$R = \frac{Fs}{S + s}$$

where:
R = BOD removal rate (lbs per day/lb MLSS)
F = Maximum BOD removal rate (lbs per day/lb MLSS)
s = Soluble effluent BOD (mg/l)
S = BOD concentration at R = 1/2 F, Michaelis Constant (mg/l)

In an extended aeration complete-mix system the BOD concentration in the aeration-surge basin is normally very low compared to the influent BOD concentration. Thus, S is usually large as compared to s. Under these conditions, the Michaelis-Menton rate can be approximated as:

$$R = \frac{Fs}{S} = ks$$

where: $k = \frac{F}{S}$ is the BOD removal coefficient expressed as lbs BOD removed per day per lb MLSS per mg/l soluble effluent BOD.

This simplification takes the form of the straight line equation with a slope k. The BOD removal per unit of MLSS is proportional to the BOD remaining in solution.

SLUDGE YIELD–Excess solids in the activated sludge system will result from the suspended solids in the influent which are not biodegraded and synthesized cell tissue which is not metabolized by endogenous respiration.

McCarty and Brodersen [10] have developed the following equation to estimate the net sludge accumulation in terms of pounds of volatile biological solids produced per pound of BOD removed:

$$\frac{A}{F} = 0.65 - \frac{0.53}{1 + \frac{E}{bM}}$$

where: A = net accumulation of volatile biological solids
F = BOD removed
E = suspended solids lost from system per day
M = total suspended solids in system
b = endogenous respiration constant

This equation assumes the net accumulation of solids to be dependent on the rate of synthesis of biological solids, and the rate of solids degradation by endogenous respiration. The endogenous respiration constant, b, is in turn dependent on the average retention time, $T_s = \frac{M}{E}$, of the biological solids (sludge age) and the temperature of the system.

NITRIFICATION–Under favorable operating conditions, oxidation of ammonia nitrogen (nitrification) can occur simultaneously with carbonaceous metabolism in activated sludge systems. The effect of nitrification on subsequent tertiary treatment of secondary effluent can be either beneficial or detrimental.

Nitrogen in the form of the ammonium ion is oxidized to the nitrate ion in a two step reaction by the autobiophic bacteria *nitrosomonas* and *nitrobacter* [11]. The reactions can be summarized as follows:

Step 1

$$2NH_4^+ + 3O_2 \xrightarrow{nitrosomonas} 2NO_2^- + 4H^+ + 2H_2O$$

Step 2

$$2NO_2^- + O_2 \xrightarrow{nitrobacter} 2NO_3^-$$

Overall Energy Reaction

$$2NH_4^+ + 4O_2 \rightarrow 2NO_3^- + 4H^+ + 2H_2O$$

In addition to being an energy source, the ammonium ion is also assimilated by the bacteria for synthesis of cell tissue. McCarty [12] represents the autotrophic synthesis reaction as follows:

Synthesis

$$4CO_2 + HCO_3^- + NH_4^+ + H_2O \rightarrow C_5H_7NO_2 + 5O_2$$

Based on both theoretical calculations and laboratory data, McCarty [12] has proposed the following overall reaction to describe the autotrophic nitrification reaction:

$$22NH_4^+ + 37O_2 + 4CO_2 + HCO_3^- \rightarrow C_5H_7NO_2 + 21NO_3^- + 20H_2O + 42H^+$$

The extended aeration activated sludge process can provide an environment well suited for nitrification. The conditions that are reported to favor nitrification include [13,14,15]:

1. Long sludge age, T_s – 5 to 20 days
2. Low organic loading – 0.15 to 0.25 lbs BOD_5/day/lb MLVSS
3. D.O. level – 1.0 to 3.0 mg/l
4. Temperature – 10 to 20 degrees C
5. Slightly alkaline pH – 7.2 to 8.0
6. Inorganic carbon source – HCO_3^-

The nitrifying autotrophic bacteria have considerably slower growth rates than the heterotrophic bacteria in the activated sludge floc. The sludge age at a given temperature must be greater than the reciprocal growth rate, if nitrification is to occur. The low organic loading improves the competitive position of the autotrophic bacteria with respect to other heterotrophic organisms.

Oxygen tension is also critical. Wuhrman [16] reports D. O. levels of 1.0 to 1.5 adequate for nitrification. Carlson [13] and others indicate D. O. concentrations of 3.0 to 4.0 mg/l, however, may be necessary for complete nitrification of available ammonia nitrogen. While a slightly alkaline pH is more favorable to nitrification, lower operational values (6.5 to 7.0) have been reported [11].

The synthesis reaction in nitrification results in a net release of hydrogen ion (H^+). Bicarbonate ion (HCO_3^-), if available, will be converted to carbonic acid (H_2CO_3) in an amount equivalent to the hydrogen ion released

$$H^+ + HCO_3^- \rightarrow H_2CO_3$$

resulting in a reduction in pH according to the equilibrium relationship:

$$pH = pK_1 + \log \frac{[HCO_3^-]}{[H_2CO_3]}$$

K_1 is the equilibrium constant for the carbonic acid – bicarbonate buffer system. At 25 degrees C, pK_1 has a value of 6.35.

Both the formation of alum floc and the efficiency of alum-phosphate precipitation are dependent upon pH. The reduction in alkalinity and consequent pH depression resulting from nitrification can be either beneficial or detrimental to subsequent tertiary treatment of secondary effluent. Formation of hydrous aluminum floc is restricted to a pH range of about 5 to 9.5 [29]. Optimum conditions for alum-phosphate precipitation occur between pH 5 to 6 [31,32].

In highly alkaline wastewater, nitrification can aid alum-phosphate precipitation and alum coagulation by reducing the pH to a more favorable level. In low alkalinity wastewater, however, the pH may be depressed below the range where these reactions can occur. The addition of soda ash or other source of alkalinity may be necessary to maintain pH control under these conditions.

Alum is also acidic. Sufficient alkalinity must be available in the secondary effluent to buffer the alum dosage required to produce the level of phosphate removal desired in the tertiary process.

TUBE CLARIFICATION

GENERAL–Tube type clarifiers were used for liquid/solids separation in both the secondary and tertiary treatment systems. Practical application of tube type clarifiers as an efficient sedimentation device is a relatively recent development in gravity separation of solids from liquids.

A thorough discussion of the fundamental principles of sedimentation has been prepared by Fair and Geyer [17]. Hansen and Culp [18] and Hansen *et al.* [19] have applied shallow depth sedimentation theory to the development of tube clarifiers. The following is a summary of sedimentation theory and the development of tube clarifiers.

SEDIMENTATION THEORY–In developing a fundamental understanding of sedimentation, Hazen [20] recognized that the removal efficiency of discrete particles settling unhindered in a basin is solely a function of (1) the settling velocity of the particles, (2) the surface area of the basin, and (3) the flow velocity through the basin. The surface area combined with the flow velocity constitutes the basin surface loading or surface overflow rate. The removal efficiency is independent of the detention time of the basin.

Camp [21] suggested the model on Figure 1 to describe unhindered settling of discrete particles in a rectangular settling basin. All particles having a settling velocity, v_s, equal or greater than $v_o = Q/A$ fall through the full depth, h_o, of the basin and are removed. Particles with a settling velocity, v_s, less than v_o, can also be removed if they are introduced at or below a height $h = v_s t_o$ above the sludge zone. The fraction of y_o particles having a settling velocity, v_s, less than v_o that are removed, y/y_o, is equal to the ratio of the velocities v_s/v_o.

$$\frac{y}{y_o} = \frac{h}{h_o} = \frac{v_s t_o}{v_o t_o} = \frac{v_s}{v_o} = \frac{v_s}{Q/A}$$

INCLINED TUBES—Sedimentation in water and wastewater treatment has conventionally been performed in large basins having detention times of 2 to 8 hours and requiring particles to settle through depths of 8 to 16 feet in order to be removed. Since sedimentation efficiency is primarily a function of surface loading, Hazen [20] suggested that an ideal settling basin should be as shallow as possible. Inserting a number of closely spaced horizontal trays in a settling basin would reduce the distance through which a particle must travel to be removed and substantially decrease the basin detention time.

Following Hazen's suggestion, several attempts [19] have been reported to utilize shallow trays or false bottoms in the design of settling basins. Flow distribution and sludge removal problems, however, limited the practicality of these designs.

To overcome the problems associated with wide, shallow trays, small diameter tubes packed parallel in bundles and inclined at various angles have been developed [18,19,22]. Longitudinal flow through tubes with diameters of a few inches optimizes hydraulic conditions for sedimentation and overcomes flow distribution problems associated with tray type settling basins. The tubes have a large wetted perimeter in relation to cross-sectional area, resulting in low Reynolds numbers and insuring laminar flow. A Reynolds number of 500 is thought to be adequate for settling [23].

As an example, a two-inch diameter tube, four feet long, operating at an equivalent surface loading of 235 gpd/sq ft has a Reynolds number of 48. The detention time is approximately three minutes at these flow conditions.

Flow distribution through parallel tubes is self-regulating. As solids deposit in a tube, the cross-sectional area decreases, resulting in increased velocities and a slightly higher head loss through the tube. If the solids accumulation is greater in one tube than another, the differential in head loss forces more flow to those tubes containing less settled solids. Thus, the flow and solids loading over the entire tube bundle is evenly distributed.

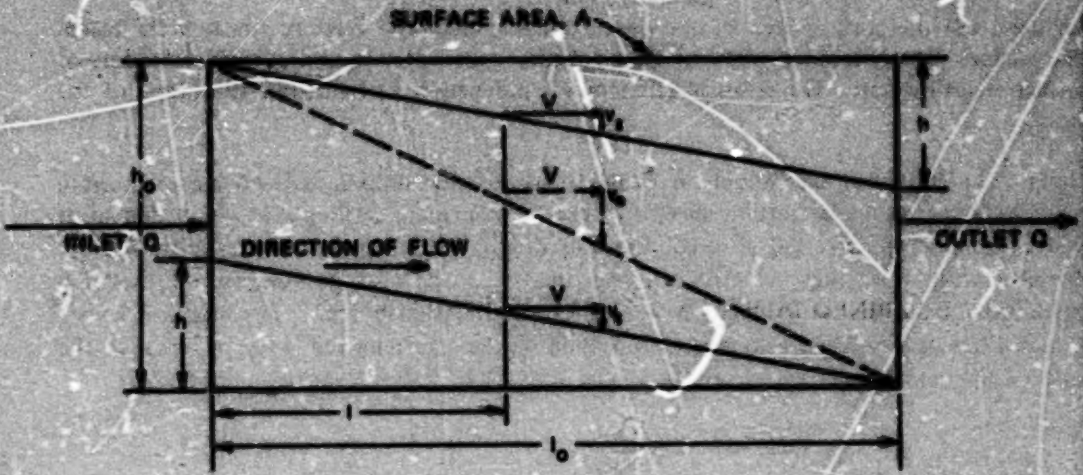

SURFACE AREA, A
V
v_s
h
h_o
V
v_o
INLET Q
DIRECTION OF FLOW
OUTLET Q
V
v_s
h
l
l_o

FIGURE 1

IDEALIZED SETTLING OF DISCRETE PARTICLES IN A HORIZONTAL FLOW BASIN

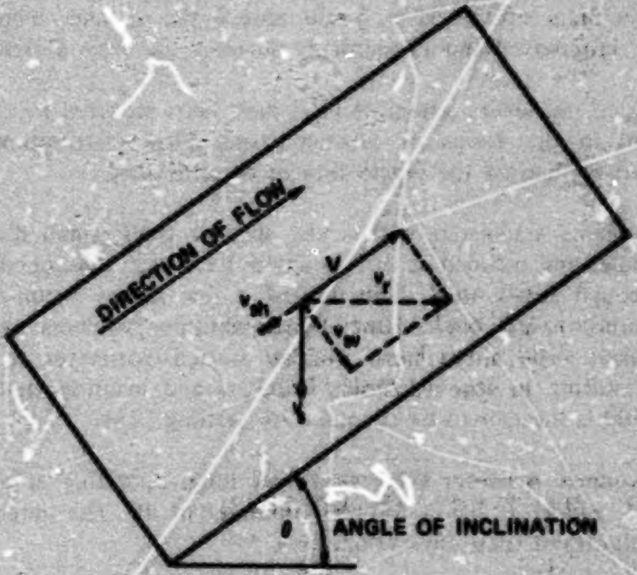

DIRECTION OF FLOW
V
v_r
v_{sh}
v_{sv}
v_s
θ
ANGLE OF INCLINATION

FIGURE 2

IDEALIZED SETTLING OF A DISCRETE PARTICLE IN AN INCLINED TUBE

Inclining the tubes in the direction of flow allows the solids that have settled to the bottom surface to slide down and eventually discharge from the tubes. As in the horizontal basin, the path traced by the particle as it settles is the vector sum of the velocity of flow, V, and the settling velocity of the particle, v_s. In inclined tubes, however, the vectors are not at right angles, as illustrated on Figure 2. The settling velocity has a component, v_{sh}, parallel and opposite to the direction of flow. If $V > v_s$, the length of settling surface decreases as the angle of inclination increases from zero up to about 25 to 30 degrees (at $V = 2.5\ v_s$,) and then increases approaching infinity at 90 degrees. For $V < v_s$, the tube length continues to decrease with increasing angle.

Experiments [18,19,22] with tubes demonstrated that an inclination of 45 to 60 degrees is necessary for continuous gravity removal of settleable material from the tubes (Figure 3a). A countercurrent flow pattern is established in which particles are carried upward as they settle, until becoming entrapped by the downward flowing stream of concentrated solids. In separating flocculant material, this countercurrent flow is thought to aid in agglomerating the smaller particles into a larger, heavier, more rapidly settling floc [19].

Tubes inclined only slightly in the direction of flow (five to eight degrees) have been successfully cleaned by periodically draining the tubes (Figure 3b). The falling water level hydraulically scours the accumulated solids from the bottom surface of the tubes.

In this research and demonstration project, both steeply and slightly inclined tubes were utilized for liquid-solids separation. Tubes inclined at 60 degrees were used in the aeration-surge basin for separation and return of mixed liquor suspended solids. Slightly inclined tubes (7-1/2 degrees) were used in the tertiary process to remove chemically coagulated and flocculated solids prior to filtration. Cleaning of the tertiary tube settler was combined with the mixed- media filter backwash operation.

TERTIARY TREATMENT

GENERAL—Tertiary treatment as applied to this project was a polishing process to remove suspended solids (turbidity) and soluble phosphate remaining in the secondary effluent. Tertiary treatment was accomplished by a four step process:

1. Alum coagulation and phosphate precipitation
2. Flocculation
3. Tube sedimentation
4. Mixed-media filtration

The theory of tube (shallow depth) sedimentation has been discussed above. A review of chemical coagulation and flocculation, aluminum-phosphate precipitation and mixed-media filtration theory is presented in the following paragraphs. More detailed discussions of chemical coagulation and precipitation have been developed by Fair and Geyer [24] and Stumm and Morgan [25]. A thorough discussion of the theory and development of mixed-media filtration has been presented by Culp and Culp [26] and Tchobanoglous [27].

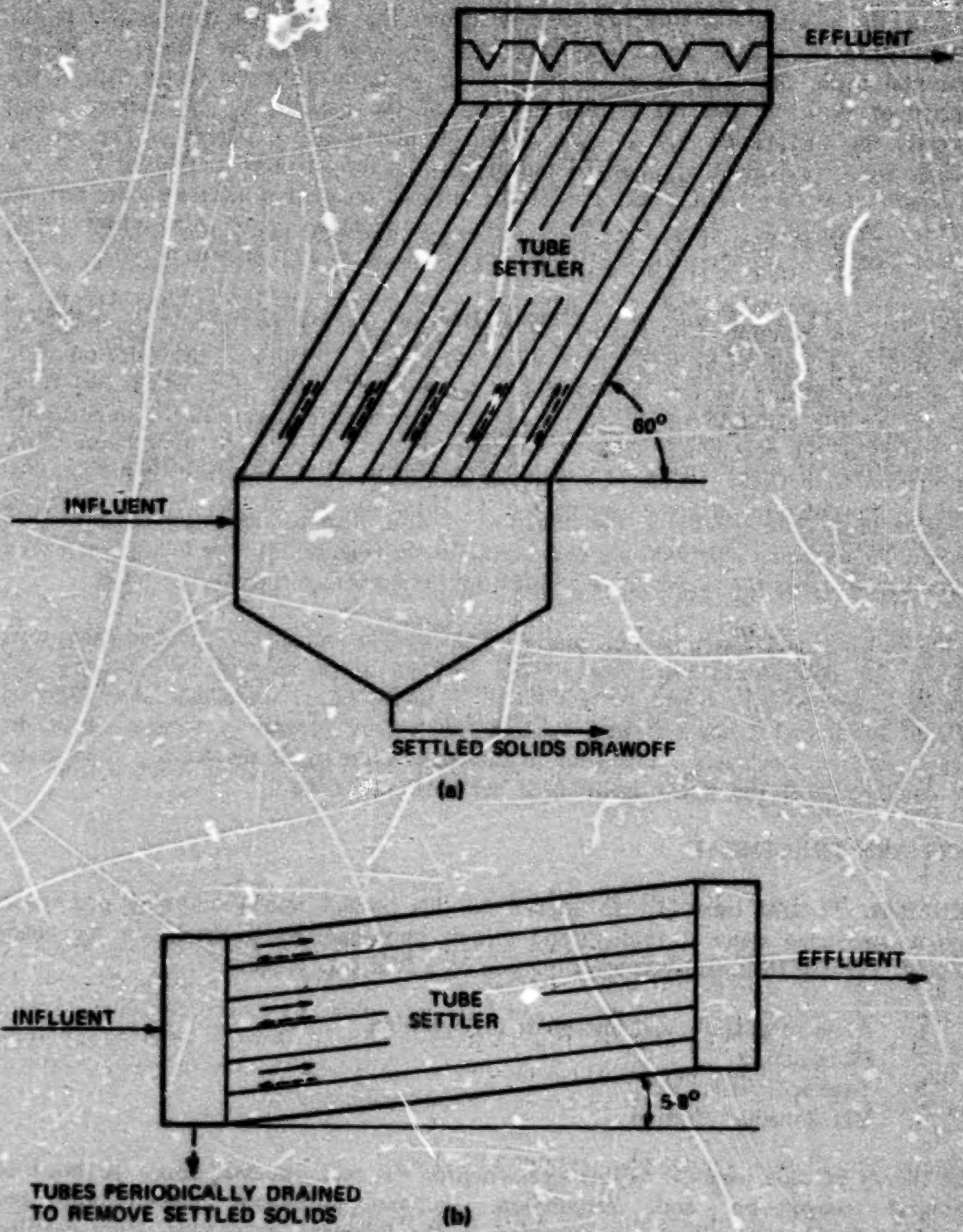

EFFLUENT
TUBE SETTLER
60°
INFLUENT
SETTLED SOLIDS DRAWOFF
(a)
EFFLUENT
TUBE SETTLER
INFLUENT
5-8°
TUBES PERIODICALLY DRAINED TO REMOVE SETTLED SOLIDS
(b)

FIGURE 3

BASIC TUBE SETTLER CONFIGURATIONS, (a) STEEPLY INCLINED TUBE SETTLER, (b) SLIGHTLY INCLINED TUBE SETTLER

CHEMICAL COAGULATION–The terms coagulation and flocculation are often used interchangeably. In this report, coagulation is considered to be the formation of complex hydrous oxides and the destabilization (or neutralization) of charged colloidal particles, resulting from the addition of floc-forming chemicals such as aluminum and iron salts. Coagulation is a complex process involving both chemical and electrostatic reactions. The reaction occurs rapidly and is essentially complete within the time necessary to physically disperse the coagulating chemicals into the liquid being treated.

Flocculation is defined as the process of agglomerating the coagulated particle into a larger mass of sufficient size to settle by gravity. This agglomeration results from the collision and sticking together or bridging of the destabilized particles. Flocculation is a time dependent reaction. Under quiescent conditions, colloidal size particles collide with each other as a result of Brownian movement. More rapidly settling particles overtake and entrap slower settling ones as the floc forms and grows. The rate of growth is very slow, however. Gentle agitation of the chemically coagulated water increases the opportunity for contact of the destabilized colloids. Gentle stirring can decrease the time required for flocculation to between 10 and 15 minutes. The addition of polyelectrolytes can also assist the flocculation process by increasing the shear strength of the floc and hastening floc formation.

A certain amount of suspended material remains in the secondary effluent after clarification. The majority of these particles are colloidal in size, ranging from 2×10^{-7} to 5×10^{-5} cm, and account for the turbidity in the water. This finely divided material consists of aggregates or single large organic molecules, such as protein, starch and tealike color compounds; cell tissue fragments; bacteria and clay particles.

The two distinguishing features of colloidal particles that affect their removal from the suspending medium are:

1. Size
2. Large surface to mass ratio or specific surface area

Colloidal particles cannot be removed by conventional filtration, but can be separated by ultrafiltration or dialysis through membranes. The surface of the particle tends to preferentially adsorb ions, usually H^+ and OH^-, becoming electrostatically charged with respect to the surrounding medium. Due to the large specific surface area, the repulsion of like charged colloids and Brownian motion are sufficient to prevent gravity settling. Under these conditions, the particles are said to be "stabilized."

Chemical coagulation is a complex equilibria involving:

1. The colloids dispersed in the water being treated
2. The pH and ionic composition of the water or dispersion medium
3. The coagulating chemical or collecting medium

The mechanism of coagulation involves:

1. Reduction of the zeta potential of the colloid

2. Neutralization of charge through the addition of oppositely charged hydrated metal oxide colloids or polyelectrolytes.

In addition, the destabilized hydrous oxides and polyelectrolytes produce gelatinous binder material to form a floc matrix during the subsequent flocculation process.

The term zeta potential is used to describe both the charge (magnitude) on the colloidal particle and the distance into the solution through which the effect of the charge extends. So long as the zeta potential exceeds a certain critical value, the colloid is stable. Below this value coagulation occurs.

The zeta potential can be reduced by:

1. Decreasing the distance of charge effectiveness
2. Neutralizing the electrostatic charge on the colloid

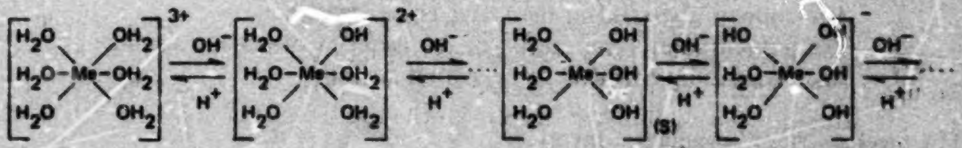
The distance through which the charge is effective is primarily dependent on the ionic concentration of the solution. The addition of di- and trivalent ions has a much greater effect on coagulation than ions of single charge. Reduction of zeta potential is brought about by the introduction of oppositely charged colloids into the suspending medium. Electrostatic attraction decreases the net charge resulting in coalescence of the colloidal particles.

Chemical coagulation of colloidal suspensions in water and wastewater can be accomplished by adding floc-forming chemicals such as alum, $Al_2(SO_4)_3 \cdot 18H_2O$; ferrous sulfate, $FeSO_4 \cdot 7H_2O$, ferric sulfate, $Fe_2(SO_4)_3$; ferric chloride $FeCl_3$ and polyelectrolytes. The initial reaction is one of solution. For alum, the reaction is:

$$Al_2(SO_4)_3 \cdot 18H_2O \rightleftharpoons 2Al^{3+} + 3SO_4^{2-} + 18H_2O$$

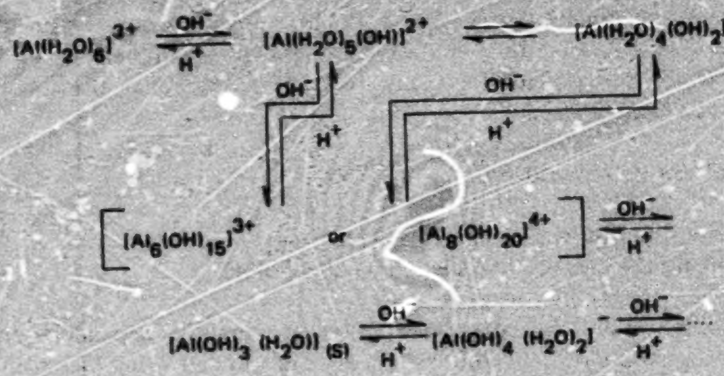
The addition of both aluminum and sulfate ions increases the ionic concentration of the medium. The net effect is a lowering of the zeta potential, creating a more favorable coagulation condition.

The second and more important reaction is hydration of the metal ion. Many metals will coordinate four to six H_2O molecules around each ion. Depending on the pH of the medium, the hydrated complex may be positively, neutrally, or negatively charged. Pfeiffer [28] in the very early 1900s visualized the hydrolysis of trivalent metal ions, such as aluminum and iron, to be represented by the equilibria:



The degree of dissociation, the species present, and the anion associated with the metal depend upon the pH and ionic composition of the solution.

Stumm and Morgan [29] have proposed a more complex model for the aluminum (III) ion.



Alum flocculation occurs in the pH range of 5 to 9.5 [29]. The hydrated aluminum complex tends to be positively charged at the lower end of the effective pH range, neutrally charged mid-range (pH, 6 to 8), and negatively charged at higher pH levels (Figure 4). Clays, bacteria, proteins, and other organic colloids present in biologically treated wastewater are generally negatively charged. Consequently, alum coagulation brought about by charge neutralization and entrapment of suspended material in the hydrous chemical floc is most effective within a pH range of 5 to 7.5.

Alum (aluminum sulfate) is a salt containing the anion (SO_4^{2-}) of a strong acid and is therefore acidic. The buffer capacity, or bicarbonate alkalinity, of the water being treated must be taken into consideration to ensure pH conditions favorable to coagulation are maintained. The following equation is a simplified illustration of the reactions that occur:

$$Al(SO_4)_3 \cdot 18H_2O + 6HCO_3^- \rightleftharpoons 3SO_4^{2-} + 2Al(OH)_3(s) + 18H_2O + 6CO_2$$

Stoichiometrically, 0.45 mg/l of bicarbonate alkalinity (expressed as $CaCO_3$) are required to react with each mg/l of alum.

Polyelectrolytes can assist both the coagulation and flocculation of finely divided suspended solids. Polyelectrolytes occur naturally as polymer derivatives of cellulose, starch, and alginates. Synthetic polyelectrolytes consist of such monomers as styrene and acrylates polymerized into high-molecular weight molecules. Depending on the functional groups (-OH, COOH, NH_2, etc.) attached to the monomers, the polyelectrolytes can be anionic, cationic, or nonionic with respect to the dissolving medium.

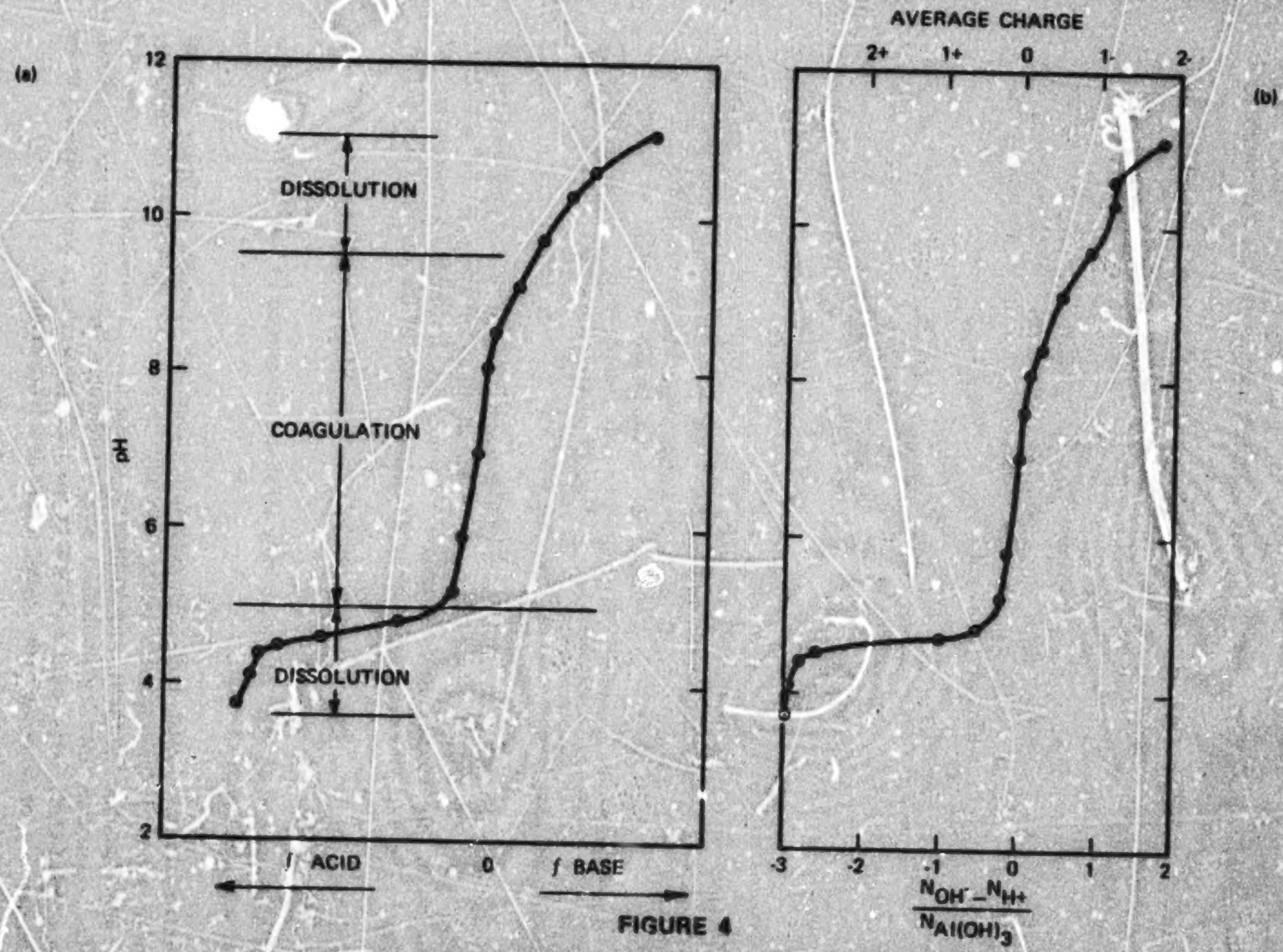


FIGURE 4

ALKALIMETRIC AND ACIDIMETRIC TITRATION OF AL(OH)$_3$ IN SOLUTION

Polyelectrolytes assist coagulation and flocculation by two general mechanisms – charge neutralization and interparticle bridging. Polymers can act as a primary coagulant, reducing the electrostatic charge of the colloidal particles. Interparticle bridging results from the polymer becoming physically attached to adsorption sites on the surface of individual particles forming a "bridge" between two or more particles. During flocculation, the bridged particles become intertwined with other bridged particles. Both the size and strength of the floc is increased, while the time required for floc formation is reduced.

PHOSPHATE PRECIPITATION–The forms of phosphorus present in domestic wastewater can be classified into three general categories: orthophosphate ion, polyphosphates or condensed phosphates, and organic phosphorus compounds. Table 1 lists some of the phosphorus species in each of these categories.

Raw domestic wastewater usually contains a considerable amount (5 to 20 mg/l as P) of all three forms. During biological treatment, however, decomposition of organic matter and hydrolysis of the condensed inorganic phosphates converts 50 to 90 percent of the total phosphate to orthophosphate [30].

The valence of condensed inorganic phosphates and orthophosphates is determined by the pH of the dissolving medium. Stumm and Morgan [25] have calculated the relative concentration of the various orthophosphate species as a function of pH from equilibrium data (Figure 5). Between pH 5 and 9, the normal range of domestic wastewaters, $H_2PO_4^-$ and HPO_4^{2-}, are the predominant species.

Phosphates can be removed from solution by addition of multivalent metal ions including Al(III), Fe(III), and Ca(II). Chemical precipitation, coagulation, and adsorption have been proposed as the removal mechanisms [24,25]. Depending upon the pH of the water being treated, one, two, or all three phenomena may be involved. The solubility relationships of the metal phosphate and metal hydroxide compounds, the hydrolysis of the metal ions, and the acid-base equilibria of the phosphate ions are all pH dependent.

Chemical precipitation of orthophosphates with Al(III) under favorable pH conditions follows the reaction:

$$Al^{3+} + H_nPO_4^{3-n} \rightleftharpoons AlPO_4(s) + nH^+$$

Stumm and Morgan [25] have calculated the total phosphate solubility for $AlPO_4(s)$ as a function of pH (Figure 6) from solubility equilibrium and acidity constants, ignoring the possible influence of phosphate-hydrous oxide complex formation. The minimum solubility at pH 6 is calculated to be 0.01 mg/l. At pH 5, the calculated solubility increases to 0.03 mg/l and at pH 7, increases to 0.3 mg/l.

As discussed earlier, the aluminum ion also undergoes a competing hydrolysis reaction to form a hydrous oxide according to the stoichiometric relationship:

TABLE I

FORMS OF PHOSPHORUS IN DOMESTIC WASTEWATER
(After 31)

FORMS	EXAMPLES
Suspended or Insoluble Organic Phosphorus	
Bacterial Cell Material	Phospholipid
Plant Debris	Phosphoprotein Nucleic Acids Polysaccharide Phosphate
Dissolved Organic Orthophosphates	
Sugar Phosphates	Glucose-1-Phosphate Adenosine Mono Phosphate
Inositol Phosphates	Inositol Mono and Hexaphosphate
Phospholipids	Glycerophosphate Phosphatidic Acid
Phosphoamides	Phosphocreatine
Orthophosphate	$H_2PO_4^-$, HPO_4^{2-}, PO_4^{3-}
Inorganic Condensed Phosphates	
Pyrophosphate	$HP_2O_7^{3-}$, P_2O^{4-}
Tripolyphosphate	$HP_3O_{10}^{4-}$, $P_3O_{10}^{5-}$
Trimetaphosphate	$HP_3O_9^{2-}$, $P_3O_9^{3-}$

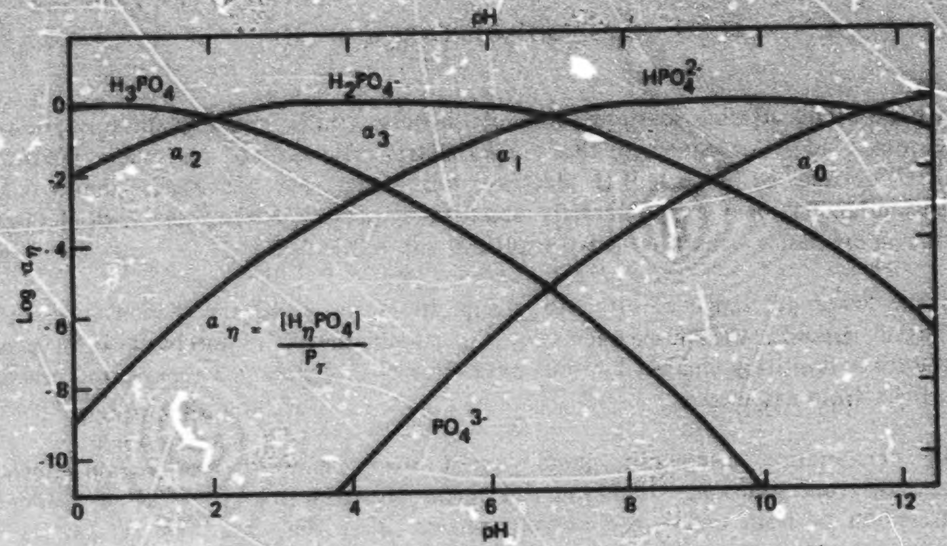


FIGURE 5

EFFECT OF pH ON VARIOUS FORMS OF ORTHOPHOSPHATE

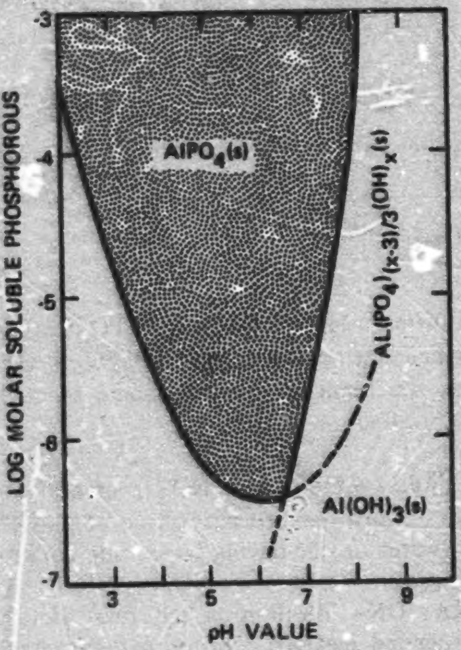


FIGURE 6

SOLUBILITY OF ALUMINUM (III) PHOSPHATE

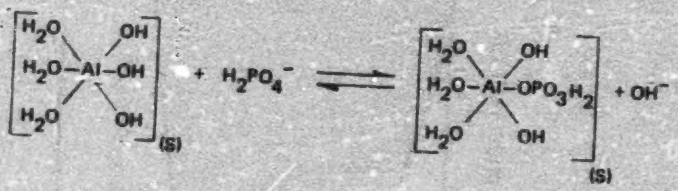
$$Al^{3+} + 6H_2O \rightleftharpoons Al(H_2O)_3(OH)_{3(s)} + 3H^+$$

In neutral and alkaline solutions, $AlPO_4(s)$ is quite soluble, favoring the formation of the hydrous oxide. Consequently, the stoichiometric efficiency of phosphate removal with Al(III) would be expected to decrease with increasing pH above about 6.0. This relationship has been experimentally observed by several investigators with the optimum removal efficiency occurring between pH 5.5 and 6.0 [31,32].

In the lower pH range of effective precipitation, the $AlPO_4(s)$ formed is a negatively charged colloid. Al(III) in excess of the stoichiometric requirement is necessary to produce a positively charged hydrous oxide floc to coagulate and remove the precipitated particles from suspension.

The precipitation mechanism alone cannot always account for the total phosphate removed from solution. Aluminum and ferric hydroxides exhibit a strong tendency for adsorption of both ortho- and polyphosphates [25]. The same type of chemical forces involved in aluminum phosphate precipitation are likely responsible for this adsorption phenomena. Phosphate adsorption onto $Al(OH)_3(s)$ generally increases as the pH decreases, tending to conform to the $AlPO_4(s)$ solubility diagram on Figure 6.

Stumm and Morgan [25] explain the mechanism of adsorption to be a replacement of an -OH group with a phosphate group in the hydrated metal oxide complex. As an example:



The OH^- has a stronger affinity for Al(III) than $H_2PO_4^-$ or HPO_4^{2-}. The extent of the exchange is, therefore, dependent on the ratio of $[H_2PO_4^-]/[OH^-]$ or $[HPO_4^{2-}]/[OH^-]$. A decrease in pH increases the tendency for phosphate to replace OH^- in the hydrated metal oxide complex or coordination sheath.

Other anions including HCO_3^-, CO_3^{2-}, and NO_3^- can also undergo a liquid exchange, competing with phosphates. Thus, reduced adsorption of phosphate would be expected in highly buffered (alkaline) water or one having large ionic concentrations.

MIXED MEDIA FILTRATION—Filtration is the final step in the tertiary treatment process for removing suspended material. The primary mechanisms involved in filtration are:

1. Mechanical straining and sedimentation
2. Physical-chemical adsorption between the suspended material and the filter media

Flocculation induced by the media and biological growth within the filter bed are secondary factors that may also contribute to particle removal [33].

Filter performance and efficiency depend to a large extent on both the characteristics of the suspended material being removed and the physical properties of the filter, including filter-media grain size, shape, density and chemical composition, and filter bed porosity. The characteristics of primary importance include:

1. Concentration
2. Particle size and distribution
3. Floc strength
4. Electrostatic charge of the particle

Of these, floc strength is commonly the dominant factor affecting solids removal efficiencies and filter performance.

Biological floc is generally much stronger than chemically induced floc. The strength of biological floc increases as the mean detention time of the solids in the biological treatment process increases. Thus, the suspended solids from an extended aeration activated sludge system would be expected to be more cohesive than that produced in contact stabilization or high rate activated sludge processes. Mechanical straining at or near the surface of the filter is the principal mechanism responsible for removal of strong floc.

Chemical floc tends to shear easily, penetrating into the filter bed. A combination of straining and adsorption is involved in separating fragile floc. The addition of polymers can add strength to the floc structure, improving particle capture in the filter bed.

Turbidity is an easily measured parameter frequently used to monitor filter performance. Tchobanoglous and Eliassen [33] have demonstrated that within limits, the suspended solids concentration can be correlated with turbidity measurements.

Rapid sand filters, using a single media with an effective particle size of between 0.35 and 1.0 mm, have conventionally been used for treatment of domestic water supplies, with limited application to filtration of biologically treated wastewater. During the backwash cycle the filter bed is hydraulically expanded to remove the entrapped particulates. The finest media rise to the top of the bed while the coarser grain sizes tend to remain at the bottom. As a result, the media is sized from fine to coarse in the normal direction of flow when filter operation is resumed (Figure 7). The effect of this gradation is for most of the suspended solids in the water being treated to be removed at or near the surface, with 75 to 90 percent of the head loss occurring in the upper inch of the filter bed [26]. The single media filter is essentially a surface filtration device. Any particles escaping the surface layer will likely pass through the bed and be present in the filter effluent.

Rapid sand filters are conventionally sized for a surface loading of 2 gpm/ft^2 and are capable of effectively treating waters with turbidities in the range of 5 to 10 Jackson Turbidity Units (JTU) [34].

To more fully utilize the full depth of the bed for removing and storing suspended solids, the concept of multimedia filtration was developed in the early 1960's. Using filter media of differing densities and particle sizes, a gradation of particle size generally from coarse to fine in the direction of flow can be achieved. Figure 8 is an idealized illustration of a mixed-media filter bed. Coarse media of low density form the upper layers while finer grain size higher density media make up the lower levels of the filter bed.

Two or three separate media with specific gravities ranging from 1.55 to 4.2 and particle sizes from 2.0 mm to 0.15 mm are commonly used. The types of media used include activated carbon (sp. gr. 1.5), anthracite coal (sp. gr. 1.65), silica sand (sp. gr. 2.6), and garnet (sp. gr. 4.2).

By carefully controlling the size distribution of each media, the average grain size will gradually decrease from the top to bottom of the filter. Distinct layers do not develop. Particles of each type of media can be found throughout the bed as illustrated on Figure 9.

The coarse to fine media distribution, or more correctly, the pore space distribution between particles, provides "in depth" filtration and storage of suspended solids throughout the entire depth of the bed. Compared to single media rapid sand filtration, more suspended material can be removed by properly designed multimedia filters before backwashing is necessary.

The decreasing void gradation of mixed-media filters allows higher filtration rates than conventional rapid sand filters. In addition, the very finely sized media (0.15 mm) provide a considerable amount of surface area for adsorption of colloidal size particles, substantially increasing filter effluent quality.

Mixed-media filters are capable of operating at hydraulic loading rates of 5 to 15 gpm and producing effluents containing one to 5 mg/l suspended solids from extended aeration activated sludge systems without chemical coagulation [26]. Chemical coagulation of secondary effluent can produce treated wastewater essentially free of suspended solids with turbidities in the order of 0.1 JTU [26].

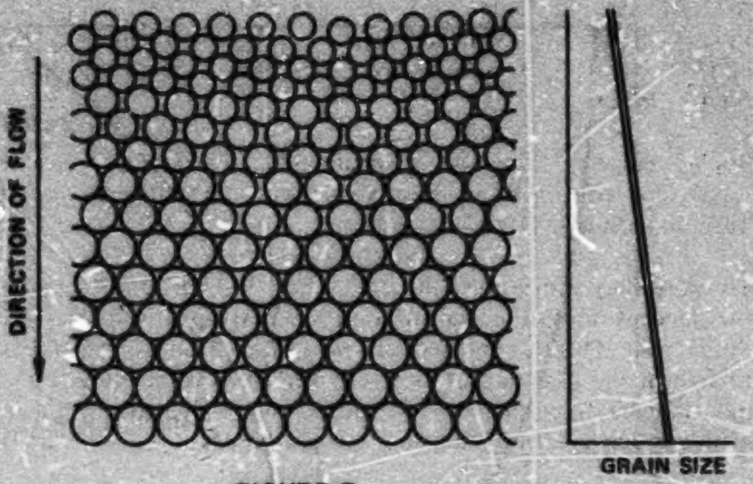

DIRECTION OF FLOW
GRAIN SIZE

FIGURE 7

CROSS-SECTION THROUGH SINGLE-MEDIA BED
SUCH AS CONVENTIONAL RAPID SAND FILTER

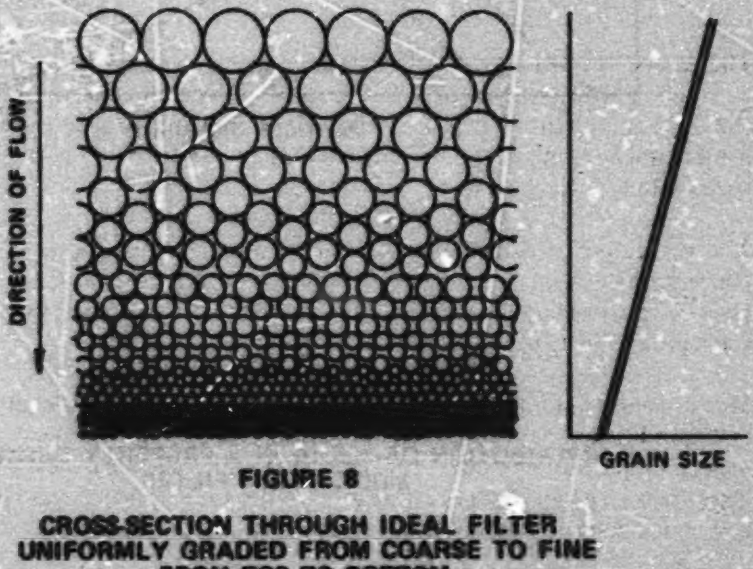

DIRECTION OF FLOW
GRAIN SIZE

FIGURE 8

CROSS-SECTION THROUGH IDEAL FILTER
UNIFORMLY GRADED FROM COARSE TO FINE
FROM TOP TO BOTTOM

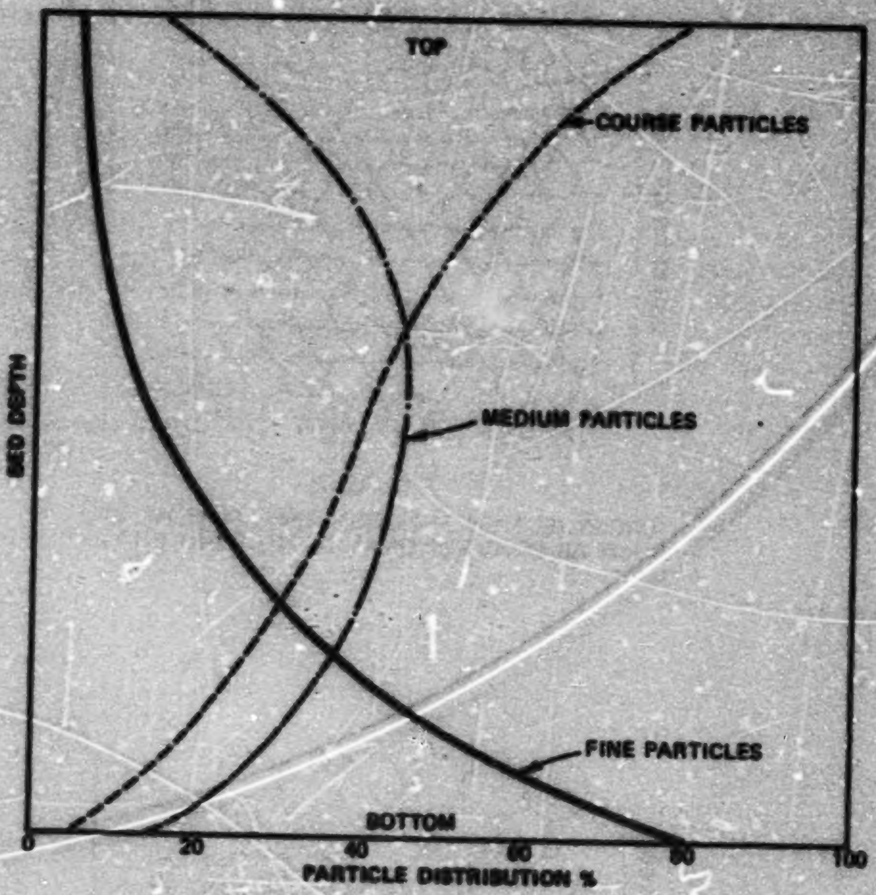


FIGURE 9

DISTRIBUTION OF MEDIA IN A PROPERLY DESIGNED MIXED-MEDIA FILTER

SECTION IV
TREATMENT FACILITIES

GENERAL DESCRIPTION

An aerial view of the Tualatin tertiary treatment plant is shown on Figure 10. The photograph was taken at the completion of construction.

Domestic waste from the City's collection system flows by gravity to the plant through a 12-inch diameter sewer to the influent pump station (1). Industrial wastewater from the Hervin Company also flows by gravity through an 8-inch diameter sewer to the influent pump station. The combined raw wastewaters are lifted by two submerged pumps through two 4-inch pipes to the headworks (2). Solids in the wastewater are normally shredded by a comminutor (3). If the comminutor is out of service for maintenance or repair, the raw influent can be bypassed through a stationary bar screen.

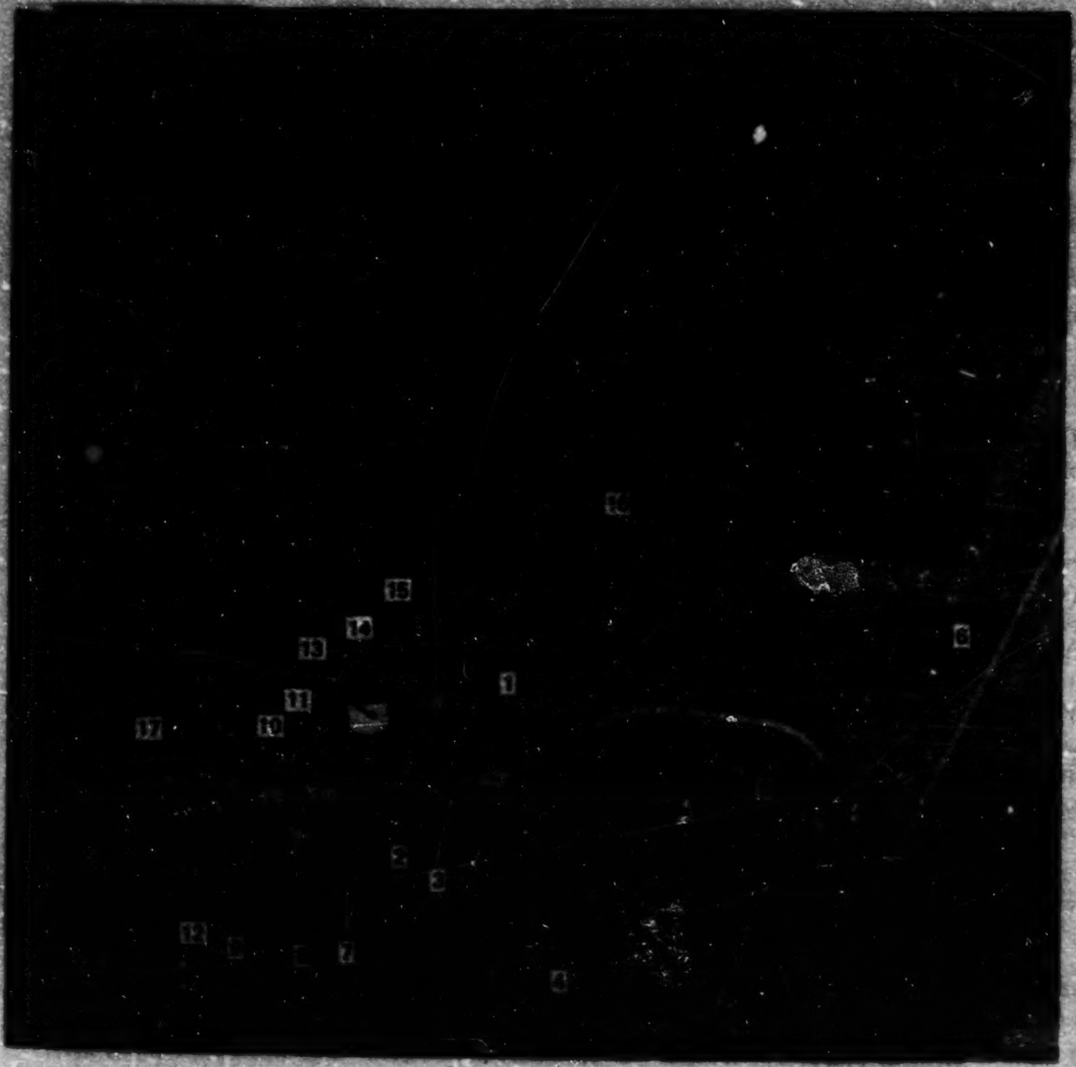
The flow from the headworks is split and continues by gravity to the aeration-surge basin (4) through two 8-inch pipelines, having outlets under each floating mechanical aerator. In addition to providing oxygen for biological treatment, the aerators evenly distribute the incoming raw waste in the basin, maintaining a completely mixed system.

The aeration-surge basin allows the tertiary portion of the plant to operate at a constant rate, as the basin water level rises and falls in response to changes in the influent flow rate. The basin effluent flows upward through the tube type secondary clarifier (5) located in the center of the aeration-surge basin, and into a collection pool above the tubes. The water level in the clarifier is essentially the same as that in the surrounding aeration-surge basin.

Activated sludge, separated from the treated wastewater in the clarifier, settles back down through the tubes. The solids are continuously drawn from the bottom of the clarifier into the aeration-surge basin by aerator induced velocity currents. Waste activated sludge is pumped through two 3-inch pipelines located directly beneath the tube clarifier to the waste activated sludge storage ponds (6). Supernatant from the sludge storage ponds is returned by gravity flow to the plant influent pump station.

Clarified secondary effluent is withdrawn at a constant rate by the tertiary influent pump (7) to an overflow box (8). The tertiary pump is automatically controlled by the water surface elevation in the aeration-surge basin. A flow splitting arrangement in the overflow box, returning a portion to the suction side of the tertiary influent pump, allows the flow rate through the tertiary system to be varied. The effluent from the secondary tube clarifier can also be discharged directly to the plant effluent pump station (14), bypassing the tertiary system.

From the overflow box, the secondary effluent continues by gravity through the flocculator tank (9), the tertiary tube settler (10), and mixed-media gravity filter (11).



1. INFLUENT PUMP STATION
2. HEADWORKS
3. COMMINUTOR
4. AERATION-SURGE BASIN
5. SECONDARY TUBE CLARIFIER
6. ACTIVATED SLUDGE PONDS
7. TERTIARY INFLUENT PUMP
8. OVERFLOW BOX
9. FLOCCULATOR TANK
10. TERTIARY TUBE SETTLER
11. MIXED MEDIA FILTER
12. ALUM STORAGE TANK
13. BACKWASH STORAGE TANK
14. EFFLUENT PUMP STATION
15. CHEMICAL SLUDGE HOLDING TANK
16. CHEMICAL SLUDGE PONDS
17. CONTROL BUILDING

FIGURE 10
TREATMENT FACILITIES

Alum is added to the tertiary influent for phosphate removal and to improve suspended solids reduction. A metering pump in the control building (17) feeds liquid alum from a storage tank (12) to the suction side of the tertiary influent pump. Provisions are made to add soda ash (sodium carbonate) on the discharge side of the tertiary influent pump for pH adjustment as necessary. Polyelectrolyte can also be added to the secondary effluent at the constant head box to aid flocculation. Both polyelectrolyte and soda ash are stored and fed through metering pumps in the control building. The flash mixing action of the tertiary pump and turbulence in the piping and overflow box disperse the chemical coagulant and flocculation aids prior to flocculation.

The flocculator inlet flow enters at the bottom of the tank. Gentle agitation by paddle blades brings the chemically coagulated solids and suspended material into intimate contact, forming floc particles of ever increasing size.

Following coagulation and flocculation the wastewater passes through an outlet near the top of the tank into the inlet of the tertiary tube settler-filter unit. The flow is split to two rows of inclined settling tubes, where the bulk of the chemically coagulated and flocculated solids are removed.

The effluent from the tube bundles collects in a center trough, discharging into the mixed media filter compartment. Remaining particulate is separated from the wastewater as the flow passes down through the filter bed. The spaces between the media grains become progressively smaller in the direction of flow. Larger particles are entrapped in the upper portion of the filter bed, while final polishing occurs in the lower levels. A gravel support and pipe lateral underdrain system collect the filtered water. To improve solids removal efficiency and control turbidity breakthrough, a small dosage of polyelectrolyte is added just prior to filtration.

Filtered wastewater is piped away through the tertiary effluent pump to the backwash storage and chlorine contact tank (13). A level controller-throttling valve system modulates the effluent flow, from the settler-filter unit to maintain a constant water level over the mixed media filter. This control system allows the effluent flow to be matched to the influent flow, regardless of changes in the influent flow rate or pressure drop across the filter as solids become trapped in the bed.

When the differential pressure across the filter rises to a preset value, the backwash cycle is automatically initiated. Both the tertiary influent and effluent pumps are stopped and the effluent valve closed. The waste valve from the tertiary tube settling compartment opens, draining the unit by gravity to the chemical sludge holding tank (15).

The receding water level carries with it the settled solids stored in the tubes, leaving the tubes clean and ready for the next filtration cycle. While the settler-filter unit is draining, the filter surface wash and backwash operations are begun.

The surface wash system consists of a fixed grid arrangement of nozzles directed downward toward the filter bed. The surface wash pump delivers water from the backwash tank to the high-pressure jets, breaking up solids deposited on the filter surface and scouring the media.

The backwash pump, also drawing water from the backwash storage tank, reverses flow through the filter underdrain system fluidizing the filter bed. A control valve maintains a constant backwash flow. As the bed expands, the suspended solids trapped by the media are carried upward out of the bed and into a wash water collection trough, discharging into the tertiary settler compartment. The relatively high upward velocity of the backwash water in the filter bed causes a rolling action of the media. The media rub against each other, tending to scour and clean the individual grains.

The initial portion of the backwash flow, containing most of the filtered suspended solids, is wasted to the tube settler compartment of the tertiary settler-filter unit scouring the bottom of the tank. Before the backwashing operation is complete, however, the waste valve is closed and the settling compartment partially refills with backwash water. The tertiary influent pump is restarted at this time. About six minutes before the backwash cycle is complete, the surface wash is discontinued to allow the media to hydraulically level. When the tube settler and filter compartments are nearly full, the backwash flow is stopped. The tertiary effluent pump is restarted and normal tertiary operation resumed.

When the backwash storage tank has refilled, the filtered wastewater overflows through an outlet near the top of the tank into the plant effluent pump station (14). Plant flow is metered in the force main from the effluent pump station. Chlorination is provided in both the backwash tank and the effluent pump station. The combined volume of the backwash storage tank and the 6-inch force main to the Tualatin River provide a minimum of one hour contact time at the average design flow to comply with the State of Oregon disinfection standards. Chlorine storage and metering facilities are located in the control building. (17).

The spent backwash water in the chemical sludge holding tank (15) is held for a predetermined period of time, allowing the heavier solids to settle. The supernatant is automatically pumped to the headworks for return to the aeration-surge basin. Settled chemical sludge is pumped from the bottom of the tank to the chemical sludge storage ponds (16). The overflow from the chemical sludge storage ponds returns by gravity through an 8-inch pipeline to the influent pump station.

An emergency bypass is provided around the plant between the influent and effluent pump stations in the event of a power outage or major failure of the secondary-tertiary system. The plant effluent can also be diverted to the chemical and activated sludge storage ponds. In the design of the plant, provisions were made for an emergency generator to provide standby power to the effluent pump station. The generator was not installed during the research and demonstration phase of the project, however.

Figure 11 is a schematic diagram of the treatment system, showing the flow pattern and location of sampling points, flow metering, main control valves and pumps.

DESIGN CONCEPT

In addition to demonstration of a tube type clarifier for separation of biological floc in an activated sludge system and chemical coagulation, flocculation, tube sedimentation, and mixed media filtration as a tertiary treatment process for extended aeration systems, the facilities were designed to minimize:

1. Construction and operating labor costs.
2. The necessity of highly trained operators.
3. Susceptibility to upset resulting from organic and hydraulic shock loads and toxic wastes.
4. Waste solids handling, stabilization, and disposal.

To accomplish these objectives, the secondary-tertiary treatment system employed the following design concepts:

Primary clarification and grit removal facilities were eliminated. Grit settles in the aeration basin and will be removed by hydraulic dredging or other means when necessary.

A plastic (PVC) membrane was used to line the aeration basin in place of higher cost reinforced concrete construction.

The complete mix extended aeration activated sludge process was selected for secondary treatment to reduce susceptibility to process upsets from organic and hydraulic shock loads and toxic wastes. Hydraulic surge capacity allowing two feet of fluctuation in water level was provided in the aeration-surge basin, permitting the tertiary system to operate at a constant rate.

Since tertiary treatment was required only during the warmer months of April through October, weather protection was not provided for the tertiary facilities, including pumps and control equipment. The majority of the tertiary piping was located above ground for ease of installation and access.

The tertiary system as well as waste sludge pumping was completely automated to minimize the necessity for operator attention.

Storage ponds were used for stabilization as well as storage of waste activated and chemical sludges, eliminating the need for separate anaerobic or aerobic sludge digestion facilities and greatly simplifying waste solids handling and disposal.

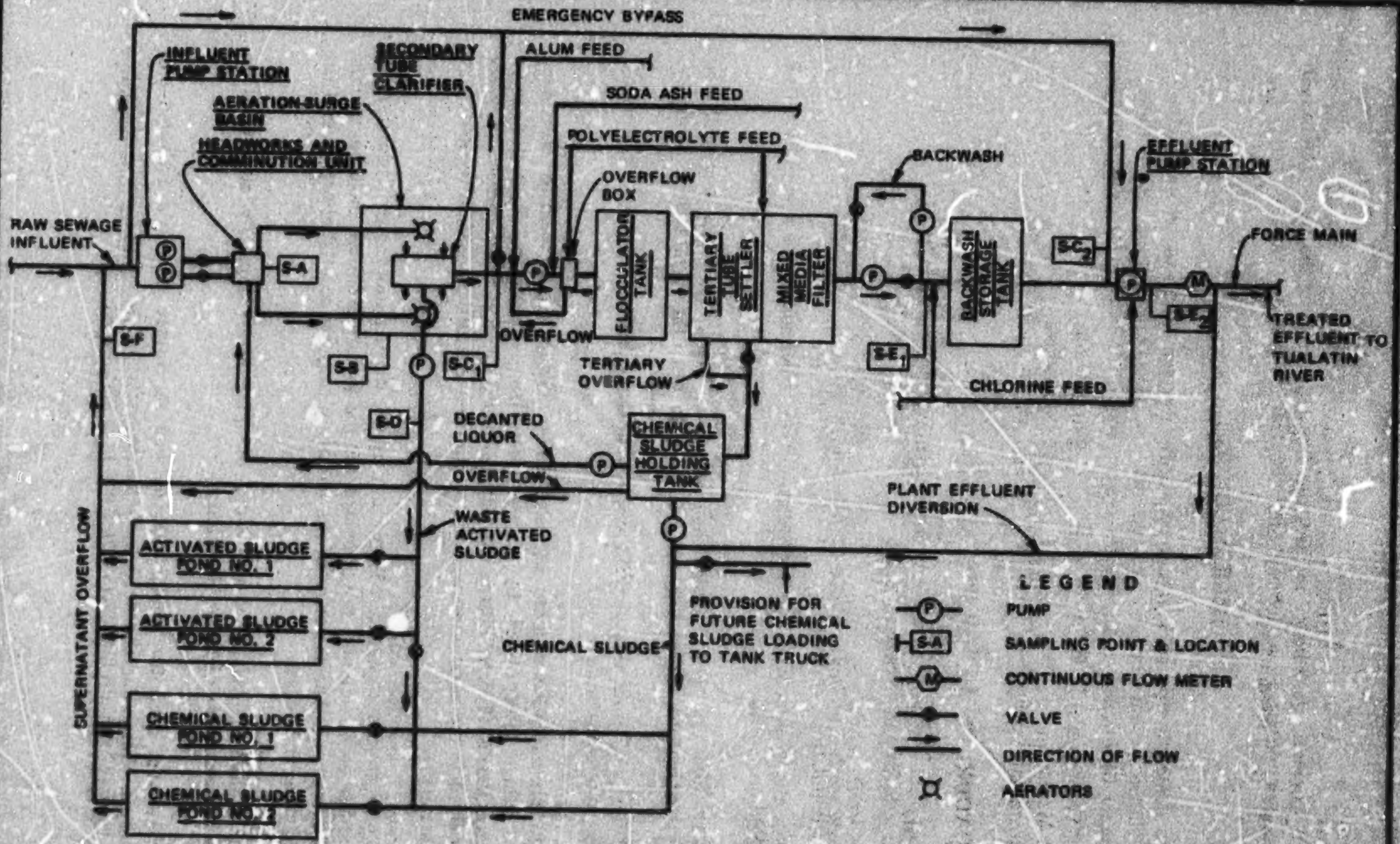

EMERGENCY BYPASS
INFLUENT PUMP STATION
SECONDARY TUBE CLARIFIER
AERATION-SURGE BASIN
HEADWORKS AND COMMINUTION UNIT
ALUM FEED
SODA ASH FEED
POLYELECTROLYTE FEED
OVERFLOW BOX
BACKWASH
EFFLUENT PUMP STATION
RAW SEWAGE INFLUENT
P
S-A
S-F
S-B
S-C1
S-D
OVERFLOW
FLOCCULATOR TANK
TERTIARY TUBE SETTLER
MIXED MEDIA FILTER
TERTIARY OVERFLOW
BACKWASH STORAGE TANK
S-C2
S-E1
S-E2
M
FORCE MAIN
TREATED EFFLUENT TO TUALATIN RIVER
CHLORINE FEED
DECANTED LIQUOR
OVERFLOW
CHEMICAL SLUDGE HOLDING TANK
WASTE ACTIVATED SLUDGE
PLANT EFFLUENT DIVERSION
SUPERNATANT OVERFLOW
ACTIVATED SLUDGE POND NO. 1
ACTIVATED SLUDGE POND NO. 2
CHEMICAL SLUDGE POND NO. 1
CHEMICAL SLUDGE POND NO. 2
CHEMICAL SLUDGE
PROVISION FOR FUTURE CHEMICAL SLUDGE LOADING TO TANK TRUCK
LEGEND
PUMP
SAMPLING POINT & LOCATION
CONTINUOUS FLOW METER
VALVE
DIRECTION OF FLOW
AERATORS

FIGURE 11

SCHEMATIC PLAN
TUALATIN TERTIARY TREATMENT SYSTEM

DESIGN CRITERIA

The criteria used for final design of the City of Tualatin tertiary treatment plant are listed in Table 2.

TABLE 2
DESIGN CRITERIA

FLOW	
Design Average	280,000 gpd
Design Maximum	636,000 gpd
Peak Instantaneous	720,000 gpd
ORGANIC LOADING	
Domestic	
BOD_5	460 lbs/day
Population Equivalent[†]	2,700
Industrial	
BOD_5	170 lbs/day
Population Equivalent	1,000
Total	
BOD_5	630 lbs/day
Population Equivalent	3,700
EFFLUENT REQUIREMENTS (TERTIARY OPERATING)	
Maximum BOD_5[††]	10 mg/l
Maximum Suspended Solids[††]	10 mg/l
Soluble Phosphate (as P)	0.1 to 1.0 mg/l
Turbidity	1 JTU

[†] Based on 0.17 lbs BOD_5/capita/day.

[††] Oregon State Department of Environmental Quality discharge standards for the Tualatin River are 10 mg/l BOD_5 and 10 mg/l suspended solids.

DESIGN FACTORS

The design factors used for the major units and equipment selection are listed in Appendix A. Photographs of selected units are included in Appendix B. Appendix C lists the manufacturers of major equipment items. The major treatment units are briefly described below.

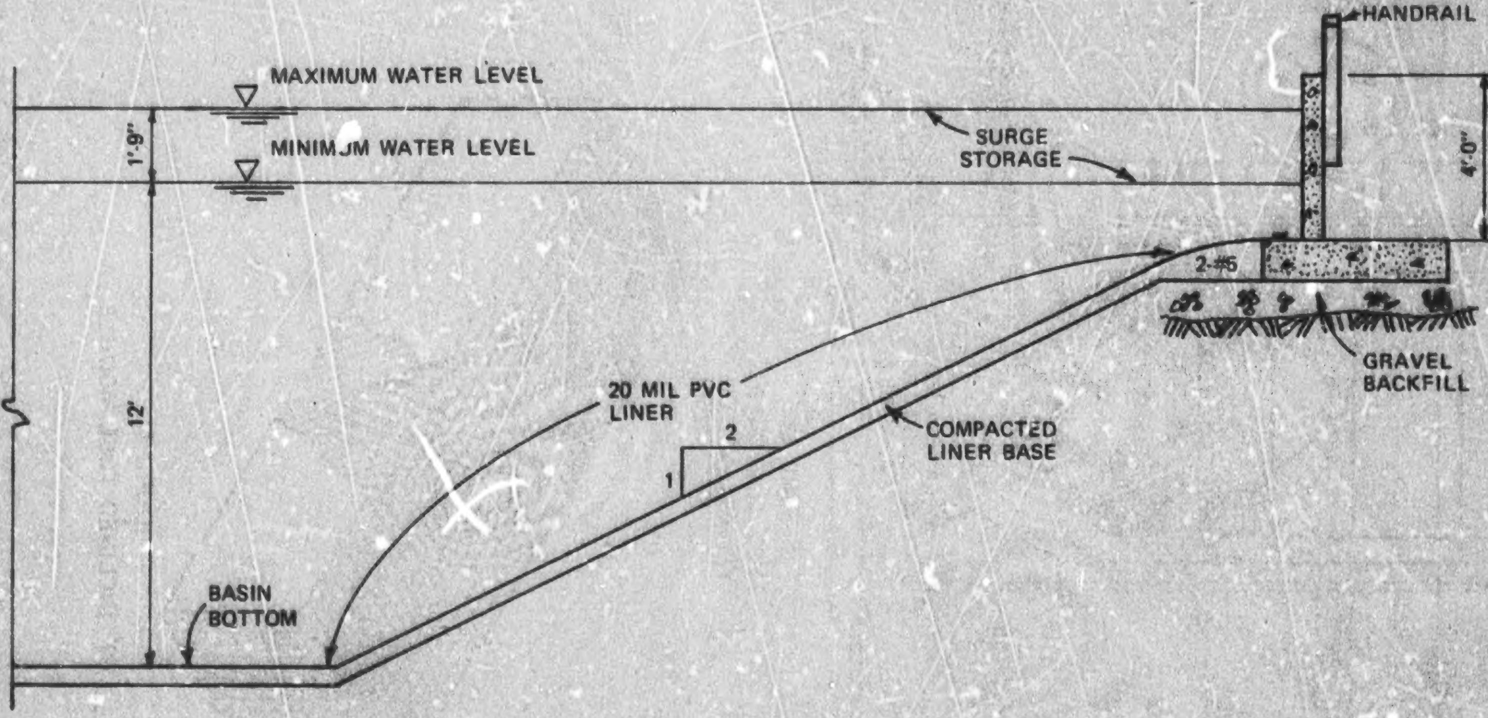
The aeration-surge basin is an earthen structure lined with 20 mil PVC plastic and surrounded by a 4-foot high concrete ring wall. A typical section through the aeration-surge basin is shown on Figure 12. The overall dimensions of the basin are approximately 70 feet by 90 feet by 12 feet minimum depth in the center portion. Two 15 hp floating mechanical aerators supply oxygen for biological treatment. The basin provides a minimum of 24 hours detention time at the design flow of 0.28 mgd, with an additional surge storage capacity of approximately 75,000 gallons. This additional storage allowance provides capacity to accommodate peak plant inflow rates in excess of the tertiary design flow rate, backwash water from the tertiary system, storage of plant flow during the backwash cycle when the tertiary unit is out of service, and supernatant returned from the chemical sludge holding tank and sludge storage ponds.

The secondary tube clarifier is located in the center of the aeration-surge basin. The unit consists of two tube bundles 6 feet wide by 20 feet long by 3-1/2 feet high, with one bundle located approximately 1-1/2 feet above the other. An end view of the tube clarifier is shown on Figure 13. Each tube bundle is made up of four separate modules fabricated by solvent welding extruded plastic channels at a 60 degree inclination from the horizontal between parallel plastic sheets. By alternating the direction of inclination of each row of channels, the module becomes a self-supporting beam that need only be supported at the ends. Each tube is 2 inches square. The plastic sheets are 0.010-inch nominal thickness rigid calendared polyvinyl chloride (PVC). The plastic channels are 0.020-inch nominal thickness acrylonitrile butadiene styrene (ABS). An illustration of a tube module resembling an egg crate is shown on Figure 14.

The tube clarifier is supported by an aluminum frame connected to a reinforced concrete slab on the bottom of the aeration basin. The lower tube module is approximately 6 feet above the basin floor. Effluent from the clarifier unit is collected in a 20-foot long, 6-inch diameter perforated collection pipe located in a collection pool above the upper tube bundle. At the design flow of 235 gpm, the tube clarifier has an overflow rate of approximately 2 gpm per square foot.

Flocculation of the secondary effluent is carried out in an 8-foot diameter by 10-foot high steel tank resting on a concrete pad. Detention time at the design flow rate is approximately 15 minutes. The flocculation mechanism consists of redwood paddles bolted to steel angle arms clamped to a solid steel drive shaft. The flocculator mechanism is mounted in a vertical position.

The tertiary tube settler and mixed-media filter are housed in a single rectangular steel tank installed on a concrete pad. Figure 15 is an illustration of the tertiary tube settler-filter unit.


HANDRAIL
MAXIMUM WATER LEVEL
MINIMUM WATER LEVEL
SURGE STORAGE
4'-0"
1'-9"
12'
2-#5
GRAVEL BACKFILL
20 MIL PVC LINER
2
1
COMPACTED LINER BASE
BASIN BOTTOM

SCALE: ¼" = 1'-0"

NOTE: INSIDE SURFACE DIMENSIONS OF BASIN 68'x88'

FIGURE 12

AERATION–SURGE BASIN
TYPICAL SECTION

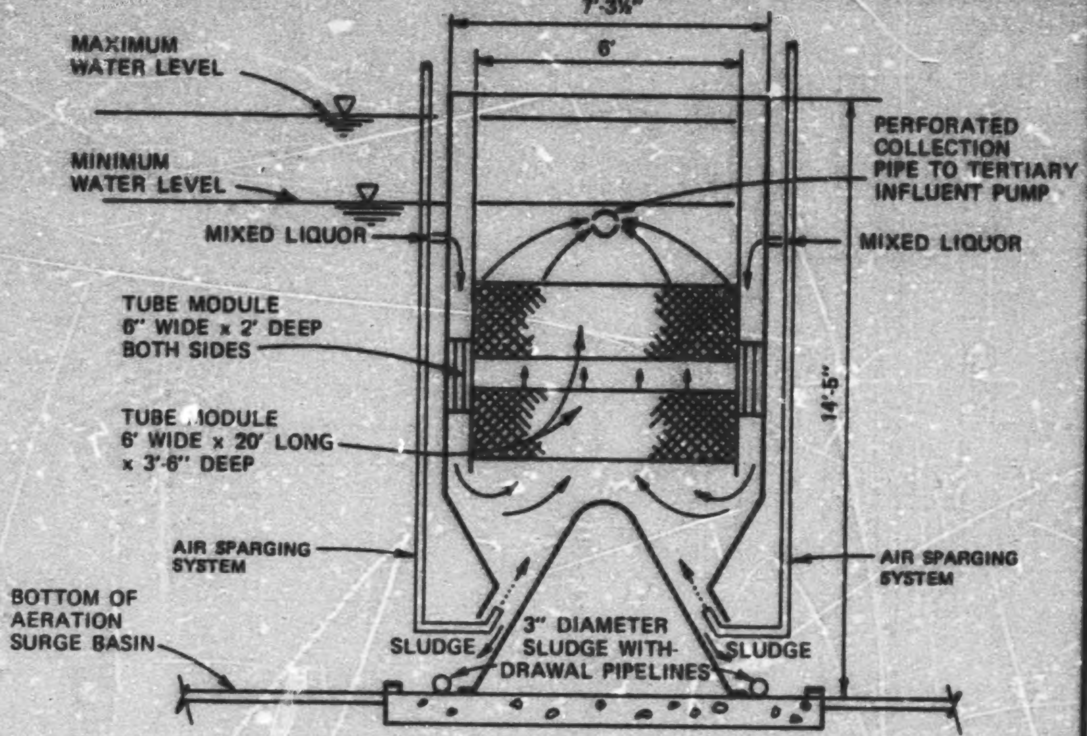

7'-3½"
6'
MAXIMUM WATER LEVEL
MINIMUM WATER LEVEL
PERFORATED COLLECTION PIPE TO TERTIARY INFLUENT PUMP
MIXED LIQUOR
MIXED LIQUOR
TUBE MODULE 6" WIDE x 2' DEEP BOTH SIDES
TUBE MODULE 6' WIDE x 20' LONG x 3'-6" DEEP
14'-5"
AIR SPARGING SYSTEM
AIR SPARGING SYSTEM
BOTTOM OF AERATION SURGE BASIN
SLUDGE
3" DIAMETER SLUDGE WITHDRAWAL PIPELINES
SLUDGE

FIGURE 13
TUBE CLARIFIER
TYPICAL SECTION

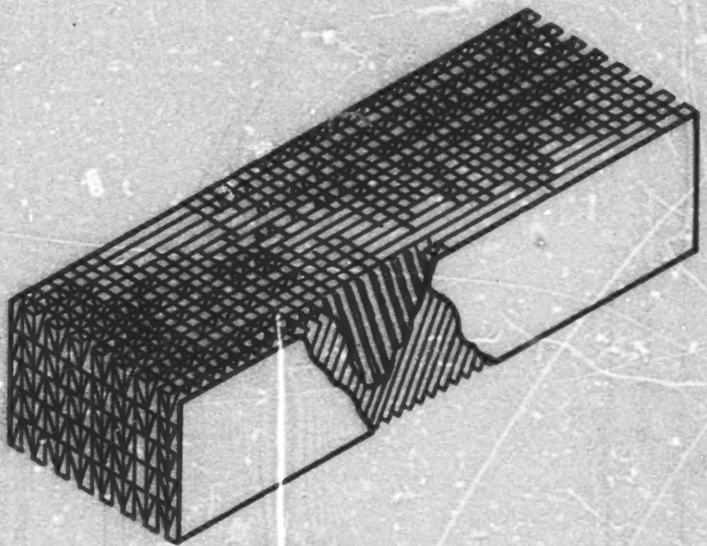
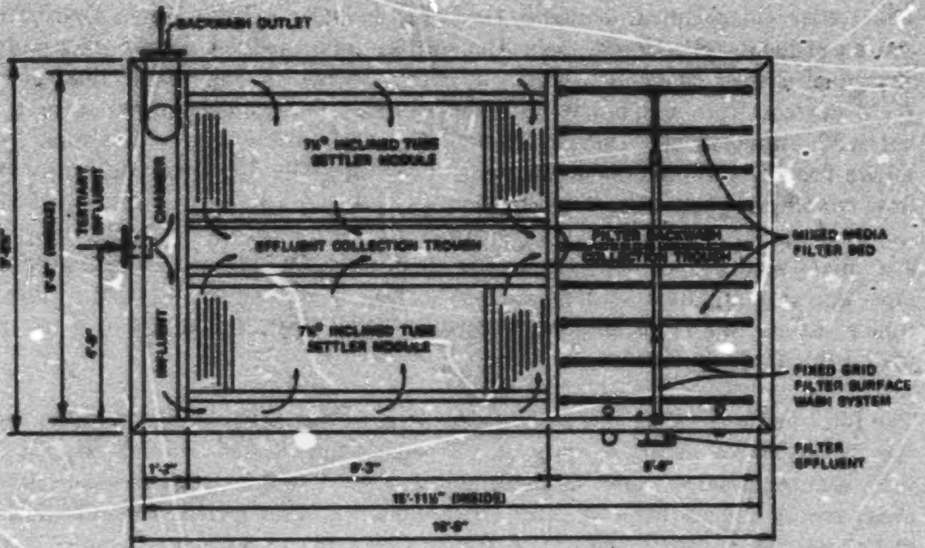
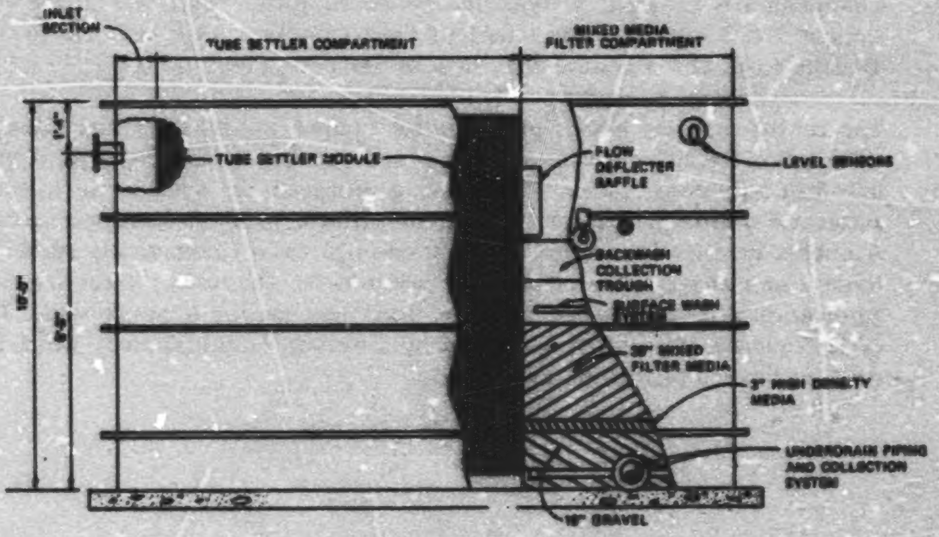


FIGURE 14
60° INCLINED TUBE MODULE


BACKWASH OUTLET
7½° INCLINED TUBE SETTLER MODULE
TERTIARY EFFLUENT
CHAMBER
EFFLUENT COLLECTION TROUGH
FILTER BACKWASH COLLECTION TROUGH
MIXED MEDIA FILTER BED
7½° INCLINED TUBE SETTLER MODULE
INFLUENT
FIXED GRID FILTER SURFACE WASH SYSTEM
FILTER EFFLUENT
7'-3" (INSIDE)
4'-9"
1'-2"
9'-3"
5'-6"
15'-11½" (INSIDE)
16'-6"

PLAN VIEW


INLET SECTION
TUBE SETTLER COMPARTMENT
MIXED MEDIA FILTER COMPARTMENT
1'-4"
TUBE SETTLER MODULE
FLOW DEFLECTER BAFFLE
LEVEL SENSORS
BACKWASH COLLECTION TROUGH
SURFACE WASH SYSTEM
30" MIXED FILTER MEDIA
3" HIGH DENSITY MEDIA
UNDERDRAIN PIPING AND COLLECTION SYSTEM
18" GRAVEL

ELEVATION VIEW

FIGURE 15

TERTIARY SETTLER/FILTER UNIT

The settler compartment contains two tube modules in addition to distribution baffles and an effluent collection trough. The settling tube modules consist of a multiplicity of hexagonal shaped channels 2 inches in depth, 39 inches long, and inclined at an angle of 7-1/2 degrees from the horizontal. The modules are fabricated from solvent welded 30 mil ABS plastic sheets. The settling unit has an overflow rate of less than 150 gpd per square foot of settling area at the design flow rate. The total basin detention time at this flow condition is approximately 30 minutes.

The mixed-media filter compartment is made up of a filter media bed, support gravel, a pipe lateral underdrain system, and a surface wash system above the filter bed. The rotary surface washarms originally installed were replaced midway through the project with a fixed grid surface wash system.

The filter bed is 5-1/2 feet wide by 9-feet long by 51 inches deep. The filter media in the top 30 inches of depth consists of 55 percent anthracite coal (1.5 sp. gr.), 30 percent silica sand (2.6 sp. gr.) and 15 percent garnet sand (4.0 sp. gr.). The particle size of the mixed-media ranges from 0.2 to 1.2 millimeters. The mixed-media is underlain by 3 inches of coarse garnet support gravel followed by 18 inches of graded stone, ranging from 3/16 to 2 inches in diameter. At the design flow, the filtration rate is approximately 4.75 gpm per square foot. The filter is backwashed at a rate of approximately 16 gpm per square foot.

Chlorination is provided in both the backwash storage tank and in the effluent pump station. The combined contact time of the backwash storage tank and the outfall to the Tualatin River at the average daily flow of 280,000 gallons per day is 65 minutes.

The control building houses plant control equipment, laboratory and office, a restroom, chlorination facilities, chemical storage and feeding equipment, and a workshop and miscellaneous storage area. The plant is designed to operate to a large degree by automatic control. The plant influent and effluent pump stations operate on water level. A bubble type level sensor in the aeration-surge basin automatically starts and stops the tertiary unit and chlorinator at preset water levels. The tertiary filter backwash system is automatically initiated, when the filter head loss reaches a preset value. The backwash cycle sequence of events is controlled by a mechanical programmer, which automatically resets at the end of each backwash cycle.

SECTION V
DEMONSTRATION PROCEDURES

PLANT STARTUP

Prior to startup, the aeration-surge basin was filled with tap water. On 1 April 1970, wastewater from the Hervin Company was turned into the plant and the aerators activated. At the same time, the aeration-surge basin was charged with two truck loads of barnyard manure and approximately 4000 gallons of activated sludge to initiate biological activity. The following week, the first domestic flow was received at the plant.

Formation of a biological floc and an increase in mixed liquor suspended solids (MLSS) were observed during the first week after startup. Activated sludge was not intentionally wasted in order to build biological solids in the basin.

The MLSS increased at a farily uniform rate from an initial value of 100 to 2200 mg/l by mid-May. During this period, the sludge volume index (SVI) dropped from 400 to 50. The secondary treatment system was assumed to be completely acclimated within approximately six weeks after startup, when BOD_5 reductions of 95 percent were achieved in the aeration-surge basin.

During the startup period, the tertiary system was operated without the addition of chemicals.

OPERATION

After startup in April 1970, routine operation and data collection for the demonstration phase of the program had to be delayed until August as a result of problems encountered at the plant. In mid-April, the plant operator/research technician suffered a serious accident and was not able to return to work full-time until mid-July. A temporary operator who had minimal laboratory experience was employed part-time to maintain and operate the facilities during the regular operator/research technician's absence. In addition, an undetected leak in the aeration-surge basin liner allowed gas generation under the membrane, causing it to tear and float. Repairs were not complete until mid-July. During this period, low influent flows allowed the plant to be operated intermittently.

Normal plant operation was resumed in late July. The tertiary system operated continuously until the last week in November, when freezing conditions damaged the chemical decant pump. During December 1970, the tertiary system was operated manually during the day. Cold weather forced suspension of tertiary operation from January through March 1971. Modifications were made to the secondary tube clarifier and tertiary equipment to improve performance while the tertiary system was out of service.

Normal tertiary operation was resumed from April 1971 to the end of October 1971. Table 3 is a detailed schedule of plant operation during the demonstration project.

DETAILED OPERATIONAL SCHEDULE

DATE	AVERAGE PLANT FLOW (GPD)	AVERAGE MLSS[1] (MG/L)	TERTIARY FLOW RATE (GPM)	CHEMICAL DOSAGE (MG/L)			
				ALUM[2]	POLYELECTROLYTE		SODA ASH[3]
					PRE-SETTLER	PRE-FILTER	
APRIL 1970	39,300	880	60,80, 160, 200	0	0	0	0
MAY	31,400	2,100	70, 80, 200	0	0	0	0
JUNE	22,300	2,100	80, 200	0	0	0	0
JULY	34,300	1,300	70, 80, 200	0	0	0	0
AUGUST	43,100	1,400	75	0, 117	0	0	0
SEPTEMBER	81,300	1,100	75, 135, 195	0, 29, 58, 88, 116-119	0	0	0 – 18.3
OCTOBER	108,100	1,100	185, 195	0, 111, 137, 160, 168, 184, 212, 238, 282	0	0	0 – 98.7
NOVEMBER	88,700	1,100	0, 188	0, 85, 158	0, 0.52, 0.62, 0.95, 1.06, 1.6, 2.0, 2.7	0	0 – 8.6
DECEMBER	112,300	1,400	0, 188	0, 96, 155, 206	0, 0.52, 0.65, 1.00, 2.0, 3.5	0	0
JANUARY 1971	118,500	1,700	0, 188	0	0	0	0
FEBRUARY	97,800	2,800	0, 120, 190	0	0	0	0
MARCH	115,100	2,400	0, 85, 165, 190	0	0	0	0
APRIL	87,900	1,900	0, 190	0, 120, 156, 210	0, 1.0, 1.4, 1.9	0, 0.10, 0.22	0 – 60.0
MAY	73,400	1,800	190	0, 156, 183	0, 0.48, 0.90, 1.00 1.3, 1.4, 1.9	0, 0.09, 0.10, 0.80	0
JUNE	94,800	2,600	190	0, 84, 126, 155, 182, 240	0, 0.45, 0.91, 1.0, 1.40, 1.80, 2.00	0, 0.08, 0.10	0
JULY	88,000	1,900	110,190	0, 63, 84, 107, 128, 155, 169, 215, 242, 285, 315	0, 1.1, 1.4, 1.5	0, 0.03, 0.04	0 – 49.0
AUGUST	118,200	2,100	110	0, 56, 81, 103-106, 135, 190, 217, 242	0, 1.1, 1.5, 1.9	0, 0.03	0
SEPTEMBER	116,000	2,800	140	108, 146	0, 0.50, 1.12	0.03, 0.04, 0.05	0
OCTOBER	107,300	2,500	140	83, 106	0	0.04	0

(1) MIXED LIQUOR SUSPENDED SOLIDS

(2) $Al_2(SO_4)_3 \cdot 14.3\ H_2O$

(3) Na_2CO_3

The research and demonstration program was officially brought to a close at the end of August 1971. However, data collected in September and October 1971 has been included in this report to expand the data base for evaluating plant operation and performance.

SAMPLING SCHEDULE AND PROCEDURES

Samples were routinely collected at various locations throughout the secondary-tertiary system. The locations of the sampling points are indicated on Figure 11 and are described below. Table 4 lists the testing program and approximate sampling schedule followed during the demonstration period.

PLANT INFLUENT

Composite samples of the plant influent were collected from the headworks at sample point A (Figure 11), located downstream of the comminutor. Samples of equal volume were composited with an automatic sampler at regular time intervals from 10 August 1970 to 15 September 1970, 22 September 1970 to 7 January 1971, and 18 May 1971 to 16 August 1971. At all other times, influent samples were manually composited from grab samples of equal volume collected 3 to 5 times during the normal daily 8-hour operating period.

AERATION-SURGE BASIN

Daily grab samples were taken directly from the aeration-surge basin, sample point B, for MLSS, SVI, and D.O. determinations.

SECONDARY EFFLUENT

Secondary effluent samples of equal volume were composited manually 3 to 5 times during the daily 8-hour operating period. When the tertiary system was operating, the secondary effluent samples were withdrawn from the tertiary influent pipeline ahead of the chemical feed locations, sample point C_1 with the aid of a hand-operated suction pump. When the tertiary system was not in service, secondary effluent samples were taken from the secondary bypass discharge to the effluent pump station, sample point C_2.

WASTE ACTIVATED SLUDGE

Grab samples were periodically taken from a tap on the discharge of the waste activated sludge pump, sample point D.

TERTIARY (PLANT) EFFLUENT

Unchlorinated tertiary effluent grab samples of equal volume were taken from a tap located in the discharge pipeline between the tertiary filter and the backwash tank, sample point E_1, and manually composited 3 to 5 times during the normal daily 8-hour period. Chlorinated tertiary (plant) effluent samples were collected periodically at the plant effluent pump station, sampling point E_2.

TABLE 4

ROUTINE SAMPLING AND TESTING SCHEDULE

(APPROXIMATE NUMBER OF TESTS PER MONTH)

	SAMPLE LOCATION	TOTAL BOD	DISSOLVED BOD	SUSPENDED SOLIDS	DISSOLVED OXYGEN	TOTAL ALKALINITY	pH	TEMPERATURE	SVI	(1) NITROGEN (TOTAL, NH_3, NO_2, NO_3)	(1) PHOSPHORUS (ORTHO, TOTAL)	Cl_2 DEMAND (BREAK POINT)	Cl_2 RESIDUAL	(2) MPN
A	PLANT INFLUENT	12		20		20	20	20		12	12			
B	AERATION-SURGE BASIN CONTENTS			20	20	20	20	20	20					
C	SECONDARY EFFLUENT	12		20		20	20	20		18	18			
D	WASTE ACTIVATED SLUDGE			P[3]										
E	TERTIARY EFFLUENT	12	12	20	20	20	20	20		18	18	P	20	P
F	SLUDGE POND SUPERNATANT	P		P			P							

[1] NITROGEN AND PHOSPHORUS, DETERMINATIONS PERFORMED ON PRESERVED SAMPLES BY EPA PACIFIC NORTHWEST ENVIRONMENTAL RESEARCH LABORATORY IN CORVALLIS, OREGON.

[2] MPN ANALYSES PERFORMED BY CITY OF PORTLAND TRYON CREEK SEWAGE TREATMENT PLANT AND PUBLIC HEALTH LABORATORY OF THE OREGON STATE HEALTH DEPARTMENT.

[3] SAMPLES TAKEN PERIODICALLY DURING DEMONSTRATION PROGRAM.

SLUDGE POND SUPERNATANT

Grab samples were periodically taken of the combined sludge pond supernatant overflow at the point of discharge into the influent pump station, sampling point F.

ANALYTICAL METHODS

All analyses, with the exception of nitrogen, phosphorus and coliform tests, were performed in the treatment plant laboratory. Nitrogen and phosphorus analyses were performed by the EPA at the Pacific Northwest Environmental Research Laboratory in Corvallis, Oregon. A limited number of coliform tests were performed by the City of Portland's Tryon Creek Sewage Treatment Plant and by the Public Health Laboratory of the Oregon State Health Department. All routine analyses were performed in accordance with *Standard Methods* [35].

NITROGEN AND PHOSPHORUS

Nitrogen and phosphorus tests were performed in accordance with the FWPCA manual [36].

A list of laboratory equipment used for the analytical testing is included in Appendix C.

SECTION VI
WASTEWATER CHARACTERISTICS

GENERAL

The domestic wastewater collection system for the City of Tualatin was constructed concurrently with the treatment plant. When the plant was started up in April 1970, a population of less than 150 was connected and contributing flow to the City sewerage system. In the initial months of the demonstration program, industrial wastewater from the Hervin Company constituted the bulk of both the flow and organic load to the plant.

Additional domestic connections were made periodically in the succeeding months of the study, increasing the contributing population to approximately 1025 by August 1971. The industrial flow, however, continued to be the major source of organic load, substantially influencing the influent wastewater characteristics and plant performance.

The industrial flow was measured and recorded beginning the last week of September 1970 through October 1971. The strength of the industrial wastewater was not continuously monitored, however. Only the combined industrial-domestic influent was sampled on a regular basis. Therefore, direct analysis of the organic loadings contributed by Hervin Company and from domestic sources was not possible.

INFLUENT CHARACTERISTICS

The following parameters of the combined industrial and domestic wastewater were measured: flow; 5-day biochemical oxygen demand (BOD_5); total suspended solids (TSS); kjeldahl, ammonia, nitrite and nitrate nitrogen; ortho and total phosphorus; total alkalinity; pH and temperature.

The data obtained during the demonstration program has been divided into five time intervals, corresponding to warm and cold weather periods and changes in the influent organic loading: (1) 3 August - 25 October 1970, (2) 26 October - 2 May 1971, (3) 3 May - 27 June 1971, (4) 28 June - 11 July 1971, and (5) 12 July - 31 October 1971. The average, maximum, minimum, and standard deviation values for the parameters measured in the plant influent during each time period and for the entire demonstration program are listed in Table 5. In the final period, nitrogen and phosphorus determinations were made during August 1971 only.

The Hervin Company normally operated on a five-day week schedule, Monday through Friday, except for a two-week vacation period in late June and early July each year. Influent samples were routinely collected during the week, but not on weekends when the treatment plant was unattended. The data in Table 5 represent the characteristics of the combined industrial and domestic flows to the plant, except for period (4). Only domestic wastewater was received during this time interval.

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TABLE 5

COMBINED INDUSTRIAL AND DOMESTIC INFLUENT WASTEWATER CHARACTERISTICS

	FLOW (MGD)	BOD_5 (MG/L)	TSS (MG/L)	KJEL-N$^{(1)}$ (MG/L)	NH_3-N (MG/L)	ORG-N (MG/L)	NO_2-N (MG/L)	NO_3-N (MG/L)	TOTAL-N (MG/L)	ORTHO-P$^{(2)}$ (MG/L)	TOTAL-P (MG/L)	ALK$^{(3)}$ (MG/L)	pH$^{(4)}$	TEMP (^{o}C)
PERIOD						3 AUGUST – 25 OCTOBER 1970								
AVERAGE	0.0739	391	397	46.9	17.8	29.0	0.05	0.21	47.1	7.56	9.25	147.	7.10	21.1
MAXIMUM	0.1524	1,300	1,300	93.0	48.0	53.0	0.44	2.40	93.1	14.3	10.3	217.	7.85	20.0
MINIMUM	0.0359	96	93	24.6	3.1	11.8	0.003	0.002	24.8	1.18	1.60	52.	6.56	16.0
STD. DEV.	0.0384	230	243	12.8	9.5	9.6	0.10	0.45	12.7	3.47	3.75	41.	0.28	2.8
PERIOD						26 OCTOBER 1970 – 2 MAY 1971								
AVERAGE	0.1030	462	472	46.2	22.5	23.7	0.06	0.11	46.3	5.53	9.45	170.	7.30	16.5
MAXIMUM	0.1520	1,170	1,640	75.0	58.0	52.7	0.32	0.80	75.0	11.0	15.0	233.	7.90	25.5
MINIMUM	0.0689	187	123	27.0	11.7	9.2	0.002	0.002	27.0	1.76	5.40	110.	6.60	10.0
STD. DEV.	0.0210	217	288	13.8	10.6	10.9	0.08	0.26	13.8	2.75	2.71	25.	0.26	3.5
PERIOD						3 MAY – 27 JUNE 1971								
AVERAGE	0.0849	640	659	64.2	19.7	49.6	0.028	0.006	65.2	6.46	12.4	188.	7.30	20.3
MAXIMUM	0.1082	987	1,330	93.0	29.0	64.0	0.052	0.008	93.0	13.0	17.9	286.	7.85	30.0
MINIMUM	0.0691	288	177	13.0	14.0	31.3	0.004	0.004	24.0	2.10	7.50	151.	5.70	14.5
STD. DEV.	0.0151	230	293	20.7	4.1	11.6	0.014	0.002	18.4	3.14	3.1	24.	0.19	3.7
PERIOD						28 JUNE – 11 JULY 1971 (DOMESTIC WASTE ONLY)								
AVERAGE	0.0746	186	294	55.3	38.9	16.5	0.05	0.024	55.4	9.90	15.8	247.	7.60	17.9
MAXIMUM	0.0778	212	400	69.3	54.0	18.5	0.12	0.086	69.3	11.0	21.0	324.	7.90	20.0
MINIMUM	0.0714	158	230	45.0	31.0	12.1	0.011	0.005	45.0	8.80	12.8	198.	7.50	16.0
STD. DEV.	0.0045	21	66	10.1	9.1	2.6	0.047	0.040	10.1	0.87	3.7	42.	0.16	1.2
PERIOD						12 JULY – 31 OCTOBER 1971$^{(5)}$								
AVERAGE	0.1101	693	618	66.8	32.8	34.0	0.016	0.023	66.8	10.1	14.2	227.	7.25	25.2
MAXIMUM	0.1225	1,520	1,730	117.	66.0	52.6	0.068	0.120	117.	17.0	20.0	359.	8.30	31.0
MINIMUM	0.0921	242	173	46.8	16.0	7.8	0.008	0.002	46.8	4.90	7.40	139.	6.75	18.5
STD. DEV.	0.0098	282	246	14.8	13.0	13.7	0.013	0.024	14.8	2.7	3.1	39.	0.31	2.7
PERIOD					TOTAL DEMONSTRATION PROJECT – 3 AUGUST 1970 – 31 OCTOBER 1971									
AVERAGE	0.0972	534	517	54.1	23.9	30.7	0.04	0.13	54.4	7.70	11.2	187.	7.25	19.8
MAXIMUM	0.1524	1,520	1,730	117.	66.0	64.0	0.44	2.40	117.	17.0	21.0	359.	8.30	31.0
MINIMUM	0.0359	96	93	13.0	3.1	7.8	0.002	0.002	24.0	1.18	1.6	52.	6.55	10.0
STD. DEV.	0.0251	276	278	17.2	12.3	13.8	0.08	0.34	16.8	3.40	4.0	47.	0.26	4.5

(1) NITROGEN REPORTED AS N (M.W. 14.01)

(2) PHOSPHORUS REPORTED AS P (M.W. 30.98)

(3) ALKALINITY REPORTED AS EQUIVALENT $CaCO_3$ (M.W. 100.09)

(4) VALUE REPORTED AS AVERAGE IS MEDIAN pH.

(5) N AND P DETERMINATION FOR JULY AND AUGUST 1971 ONLY, TEMPERATURE DETERMINED FOR JULY, AUGUST AND SEPTEMBER ONLY.

The Hervin Company suspended operations for two weeks each summer for annual employee vacations. During these periods in both 1970 and 1971 influent samples were collected to characterize the domestic contribution to the influent wastewater in the absence of infiltration and to provide data to estimate BOD, nitrogen and phosphorous loadings during the weekends. The results of these analyses, together with data from four samples collected on other days, when only domestic wastewater was received by the plant, are summarized in Table 6.

FLOW

Daily plant flow was measured in the effluent from the treatment plant. Due to the surge capacity in the aeration-surge basin and the low flows experienced during the demonstration period, the daily influent and effluent flows were not necessarily equivalent.

The daily effluent flow averaged over the week, however, was considered to be representative of the average daily influent flow.

The average flow during the demonstration program was about 0.10 mgd. The average weekly flow varied from a minimum of 0.0359 mgd to a maximum of .1524 mgd. Figure 16 is a plot of the average weekly flow, rainfall, and population contributing to the plant versus time for the demonstration period. The effect of infiltration during wet weather periods, as well as the general increase in flow with population, is evident.

BIOCHEMICAL OXYGEN DEMAND

The BOD_5 of the combined industrial-domestic weekday influent flow averaged 534 mg/l and varied between a minimum of 96 mg/l and a maximum of 1520 mg/l during the demonstration period. The BOD_5 of the domestic flow, determined from influent samples collected when Hervin Company was not operating and infiltration was not a factor, averaged 230 mg/l. Figure 17 is a plot of the average weekday BOD_5 concentration versus time.

Also shown on Figure 17 is the 7-day average weekly BOD_5 concentration, weighted to include the weekend domestic flow. The 7-day average weekly BOD_5 concentration was calculated from the following equation:

$$\text{7-day average BOD (mg/l)} = \frac{[\text{5 day } \overline{\text{BOD}} \times \overline{Q}_5 \times 8.34] \times 5 + [\overline{\text{POP}} \times \text{P.E.}] \times 2}{\overline{Q}_7 \times 8.34 \times 7}$$

TABLE 6

DOMESTIC WASTEWATER INFLUENT CHARACTERISTICS

PARAMETER	AVERAGE	RANGE
BOD (MG/L)	230.	158. – 355.
TSS (MG/L)	275.	148. – 512.
KJEL-N (MG/L)	54.1	45.0 – 69.3
NH_3-N (MG/L)	38.9	31.0 – 54.0
ORG-N (MG/L)	15.2	12.1 – 18.5
NO_2-N (MG/L)	0.051	0.011 – 0.12
NO_3-N (MG/L)	0.04	0.005 – 0.12
TOTAL-N (MG/L)	54.2	45.0 – 69.3
ORTHO-P (MG/L)	11.1	8.8 – 17.0
TOTAL -P (MG/L)	16.5	12.8 – 20.0

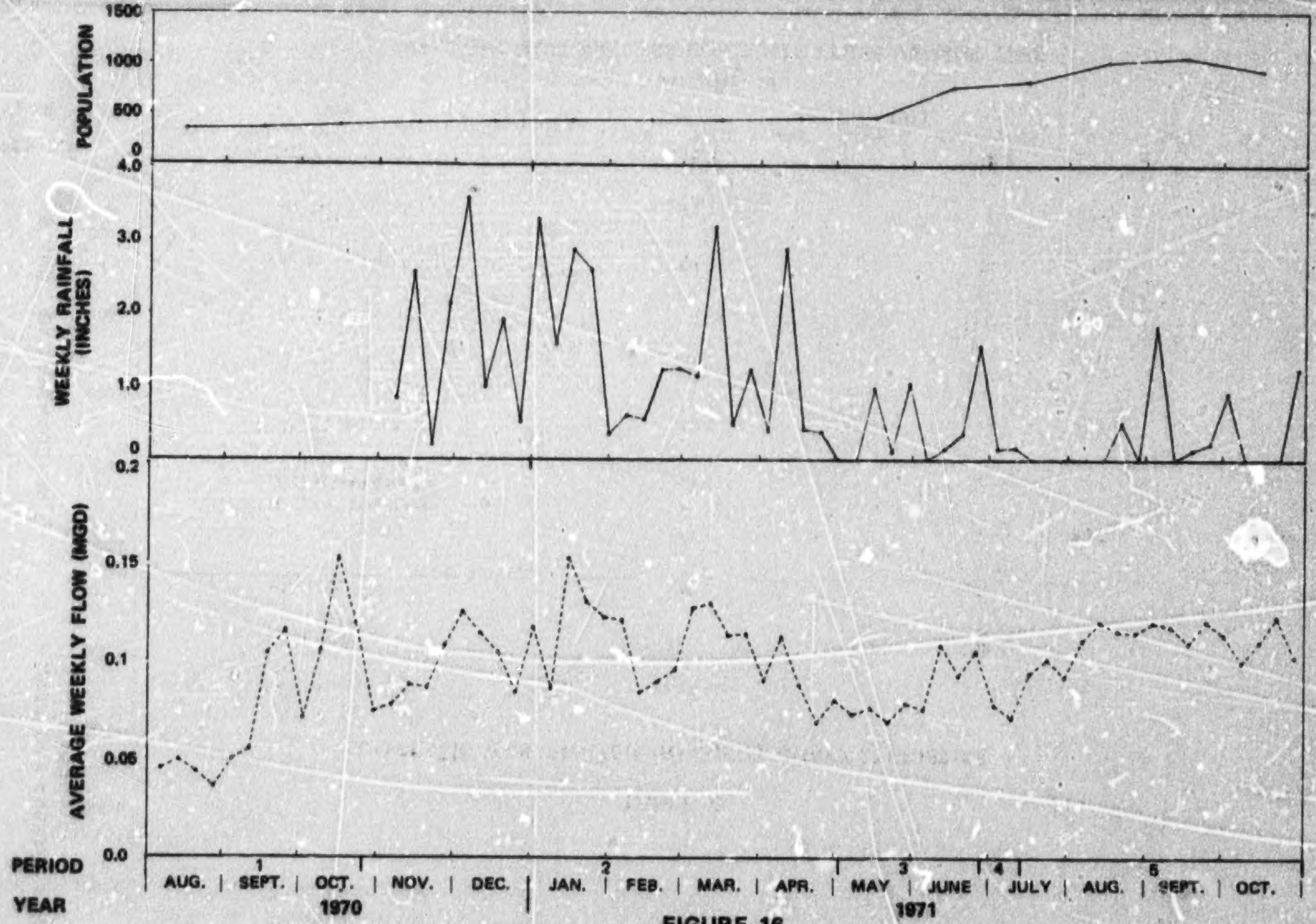


FIGURE 16

POPULATION, RAINFALL AND FLOW VERSUS TIME

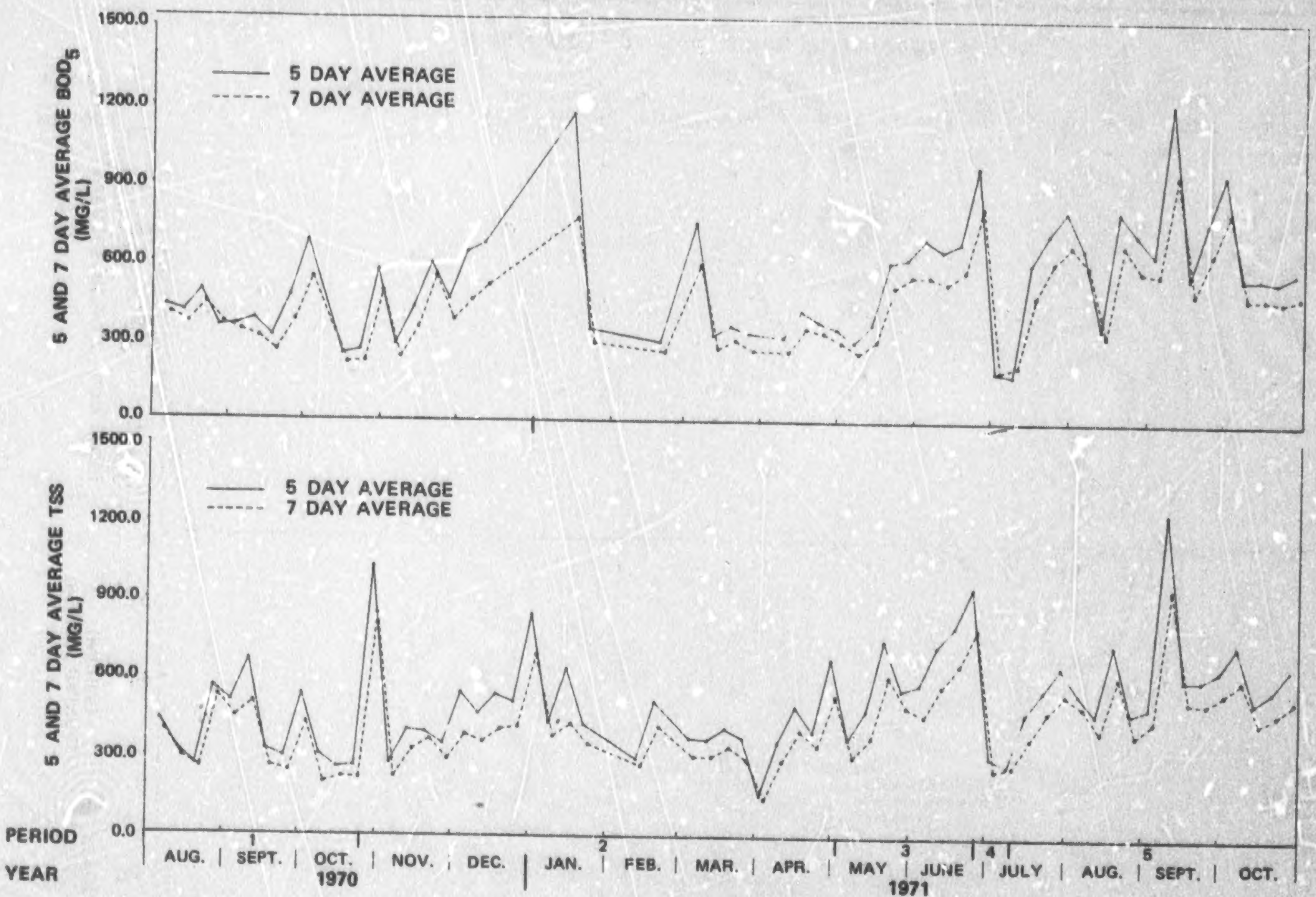


FIGURE 17

INFLUENT BOD AND TSS CONCENTRATION VERSUS TIME

Where: 5 day $\overline{BOD}$ = Average weekday BOD_5 concentration (mg/l)
$\overline{Q}_5$ = Average weekday flow (mgd)
$\overline{Q}_7$ = Average 7-day weekly flow (mgd)
$\overline{POP}$ = Average monthly population connected to the plant
P.E. = Calculated population equivalent, or per capita, BOD loading (lbs/cap - d)
8.34 = Conversion constant ([lbs/10^6 gal]) / [mg/l])

This equation was also used to calculate 7-day weekly averages for influent TSS.

The per capita loadings were developed from the domestic wastewater characterization data in Table 6 and are discussed in more detail later in this section.

TOTAL SUSPENDED SOLIDS

The TSS concentration averaged 517 mg/l for the combined industrial and domestic influent flow over the demonstration period and ranged between a minimum of 93 mg/l to a maximum of 1730 mg/l. The average TSS concentration determined for the domestic flow only was 275 mg/l. The average weekday TSS concentration and 7-day average weekly TSS concentration are also shown on Figure 17.

NITROGEN

The influent nitrogen was present mainly as organic nitrogen and ammonium ion. The total nitrogen content of the combined industrial and domestic influent averaged 54.4 mg/l (as N) during the project, varying from a minimum of 24.0 mg/l to a maximum of 117 mg/l. The average nitrite plus nitrate nitrogen for the combined influent was 0.17 mg/l. The total nitrogen content of the domestic only contribution averaged 54.2 mg/l with nitrite and nitrate nitrogen constituting less than 0.1 mg/l.

PHOSPHORUS

The ortho and total phosphorus concentrations of the combined influent averaged 7.7 and 11.2 mg/l (as P), respectively. The orthophosphate levels ranged between 1.18 and 17.0 mg/l and the total phosphate concentrations between 1.60 and 21.0 mg/l. The average ortho and total phosphorus contents for domestic flow only were determined to be 11.1 and 16.5 mg/l, respectively.

NUTRIENT RATIO

The average BOD_5:nitrogen(as N):phosphorus (as P) ratio in the combined industrial and domestic influent for the demonstration period was 100:10.2:2.1. The average nutrient ratio of the domestic waste contribution was 100:23.6:7.2.

TOTAL ALKALINITY AND pH

The total alkalinity tended to increase during the course of the project from an average value of 148 mg (as $CaCO_3$) during August-October 1970 to 235 mg/l in the period August-October 1971. The average value of the alkalinity was 187 mg and ranged from 52 to 359 mg/l. The pH ranged between 6.55 and 8.30. The median pH value of the influent was 7.25.

TEMPERATURE

The influent temperature ranged from 10 to 31 degrees C and averaged 20 degrees C.

LOADING

The daily flow (mgd) and BOD_5 TSS, total nitrogen and total phosphorous loadings (lbs per day) of the combined industrial and domestic influent were averaged weekly for the five week days (Monday - Friday). Average daily flows and loading were also calculated on a 7-day week basis, to include the effect of domestic flow only on weekends. The seven-day week average daily loadings were calculated using the following equation:

7-day week average daily loading (lbs/day) =

$$\frac{[\text{Average weekday loading (lbs/day)}] \times 5 + [\overline{\text{POP}} \times \text{P.E.}] \times 2}{7}$$

Where: $\overline{\text{POP}}$ = Average monthly population connected to the plant

P.E. = Calculated population equivalent (lbs/cap - d)

The average, maximum, minimum, and standard deviations for each of the loading parameters are summarized in Table 7 for the five separate time periods and for the total duration of the demonstration project.

FLOW

The average weekday flow for the demonstration period was 0.1019 mgd and varied from a minimum of .0419 during the week of 24-28 August 1970 to a maximum of .1704 mgd during the week of 12-16 October 1970. Over the 7-day week, the daily flow averaged 0.0972 mgd for the demonstration period. The minimum and maximum 7-day week average daily flows of 0.0359 mgd and 0.1524 mgd occurred during the same respective weeks also.

BOD

The average weekday BOD_5 loading ranged from 110 lbs per day during the week of 5-9 July when Hervin Company was shut down for annual employee vacations, to a maximum of 1,340 lbs per day in the week of 11-15 January 1971. The average weekday

TABLE 7

INFLUENT LOADINGS

		5-DAY WEEK (MONDAY–FRIDAY)				7-DAY WEEK (MONDAY–SUNDAY)			
		AVERAGE	MAXIMUM	MINIMUM	STD. DEV.	AVERAGE	MAXIMUM	MINIMUM	STD. DEV.
PERIOD		3 AUGUST – 25 OCTOBER 1970							
FLOW	(MGD)	0.0822	0.1704	0.0419	0.0408	0.0788	0.1524	0.0350	0.0304
BOD$_5$	(LBS/DAY)	280.	481.	126.	117.	207.	386.	110.	84.
TSS	(LBS/DAY)	247.	370	106.	78.6	195.	284.	94.	57.
TOTAL–N	(LBS/DAY)	30.9	51.6	17.9	6.9	26.5	41.5	17.0	7.2
TOTAL–P	(LBS/DAY)	5.6	8.2	2.5	1.8	5.3	8.0	3.2	1.3
PERIOD		26 OCTOBER 1970–2 MAY 1971							
FLOW	(MGD)	0.1081	0.1410	0.0738	0.0204	0.1030	0.1520	0.0689	0.0210
BOD$_5$	(LBS/DAY)	456.	1340.	184.	278.	362.	963.	157.	168.
TSS	(LBS/DAY)	420.	916.	121.	175.	322.	676.	110.	125.
TOTAL–N	(LBS/DAY)	40.1	81.1	20.8	14.8	33.8	63.0	20.3	10.5
TOTAL–P	(LBS/DAY)	8.4	15.6	4.6	2.9	7.6	12.7	4.9	2.1
PERIOD		3 MAY–27 JUNE 1971							
FLOW	(MGD)	0.0893	0.1159	0.0732	0.0171	0.0848	0.1082	0.0691	0.0151
BOD$_5$	(LBS/DAY)	478.	943.	186.	237.	377.	721.	161.	177.
TSS	(LBS/DAY)	507.	934.	229.	220.	282.	890.	186.	171.
TOTAL–N	(LBS/DAY)	47.1	77.9	15.5	21.6	40.8	65.0	16.7	17.2
TOTAL–P	(LBS/DAY)	9.4	15.9	4.7	3.8	8.9	14.2	5.1	3.2
PERIOD		28 JUNE – 11 JULY 1971 (DOMESTIC WASTE ONLY)							
FLOW	(MGD)	0.0720	0.0738	0.0702	0.0025	0.0746	0.0778	0.0714	0.0045
BOD$_5$	(LBS/DAY)	111.	113.	110.	2.	128.	128.	128.	0.
TSS	(LBS/DAY)	179.	181.	170.	8.	166.	169.	163.	4.
TOTAL–N	(LBS/DAY)	32.3	35.9	28.6	5.1	32.6	34.9	30.3	3.2
TOTAL–P	(LBS/DAY)	9.7	10.6	8.7	1.4	9.8	10.6	9.1	1.1
PERIOD		12 JULY – 31 OCTOBER 1971							
FLOW	(MGD)	0.1161	0.1323	0.0905	0.0122	0.1101	0.1225	0.0921	0.0096
BOD$_5$	(LBS/DAY)	670.	1211.	390.	212.	538.	930.	333.	154.
TSS	(LBS/DAY)	607.	1235.	357.	192.	483.	937.	297.	138.
TOTAL–N	(LBS/DAY)	63.3	75.4	43.8	11.2	56.6	86.1	41.1	8.8
TOTAL–P	(LBS/DAY)	12.9	15.9	9.5	1.8	12.7	15.1	10.5	1.3
PERIOD		TOTAL DEMONSTRATION PROJECT – 3 AUGUST 1970 – 31 OCTOBER 1971							
FLOW	(MGD)	0.1019	0.1704	0.0419	0.0266	0.0972	0.1524	0.0350	0.0251
BOD$_5$	(LBS/DAY)	470.	1340.	110.	269.	373.	963.	110.	201.
TSS	(LBS/DAY)	438.	1235.	106.	213.	343.	937.	94.	160.
TOTAL–N	(LBS/DAY)	42.6	81.1	15.5	17.7	37.1	86.1	16.7	14.6
TOTAL–P	(LBS/DAY)	8.7	15.9	2.5	3.5	8.2	15.1	3.2	3.2

BOD_5 loading for the demonstration period was 470 lbs per day. The absence of industrial flows on weekends reduced the daily loads averaged over a 7-day period by about 20%. Daily BOD loadings averaged over the 5-day and 7-day week are plotted versus time on Figure 18.

The treatment plant experienced organic overload conditions (design capacity 630 lbs BOD per day) during one or more weeks in all but the first three-month period. During August through October 1971, the average weekday BOD loading was 689 lbs per day. As a result, satisfactory D.O. levels could not be maintained in the aeration-surge basin. Secondary treatment performance was impaired, particularly during the later phases of the demonstration program. The effect of organic overloading on plant performance and operation is discussed in Sections VII and VIII.

TSS

The average weekday TSS loading ranged from 106 lbs per day during the week of 17-21 August 1970 to 1235 lbs per day during the week of 6-10 September 1971. The weekday TSS loading averaged 438 lbs per day for the demonstration period. When averaged over the 7-day week, the domestic weekend flow reduced the weekday TSS load by approximately 22 percent. The 5-day and 7-day week average daily TSS loadings are shown on Figure 18.

TOTAL NITROGEN

The weekday total nitrogen loading averaged 42.6 lbs per day over the total demonstration period and varied from a minimum of 15.5 lbs per day during the week of 31 May-4 June 1971 to a maximum of 81.1 lbs per day during the week of 28 December 1970-1 January 1971. Averaging the total nitrogen loading over 7 days reduced the average weekday loading approximately 12.5 percent. The 5-day and 7-day week average daily nitrogen loadings are plotted versus time on Figure 19.

TOTAL PHOSPHORUS

The weekday total phosphorus loadings ranged from 2.5 lbs per day during the week of 19-23 October 1970 to 15.9 lbs per day during the weeks of 21-25 June 1971 and 2-8 August 1971. The average weekday total phosphorus loading for the demonstration period was 8.7 lbs per day. Averaging the total phosphorus loading over a 7-day week reduced the 5-day average loading by approximately 4.5 percent. The 5-day and 7-day week average daily total phosphorus loadings are plotted versus time on Figure 19.

INDUSTRIAL WASTEWATER

Both the flow and BOD of the industrial wastewater from Hervin Company varied widely. The daily flows ranged from 1,600 to 101,800 gpd. The average daily flow from the last week of September 1970 through August 1971 was 26,400 gpd. BOD analyses run on industrial wastewater samples during and after the demonstration period ranged between 840 and 3400 mg/l. It is estimated that between 50 and 65 percent of the organic load, applied to the treatment plant during the demonstration program, resulted from industrial wastewater.

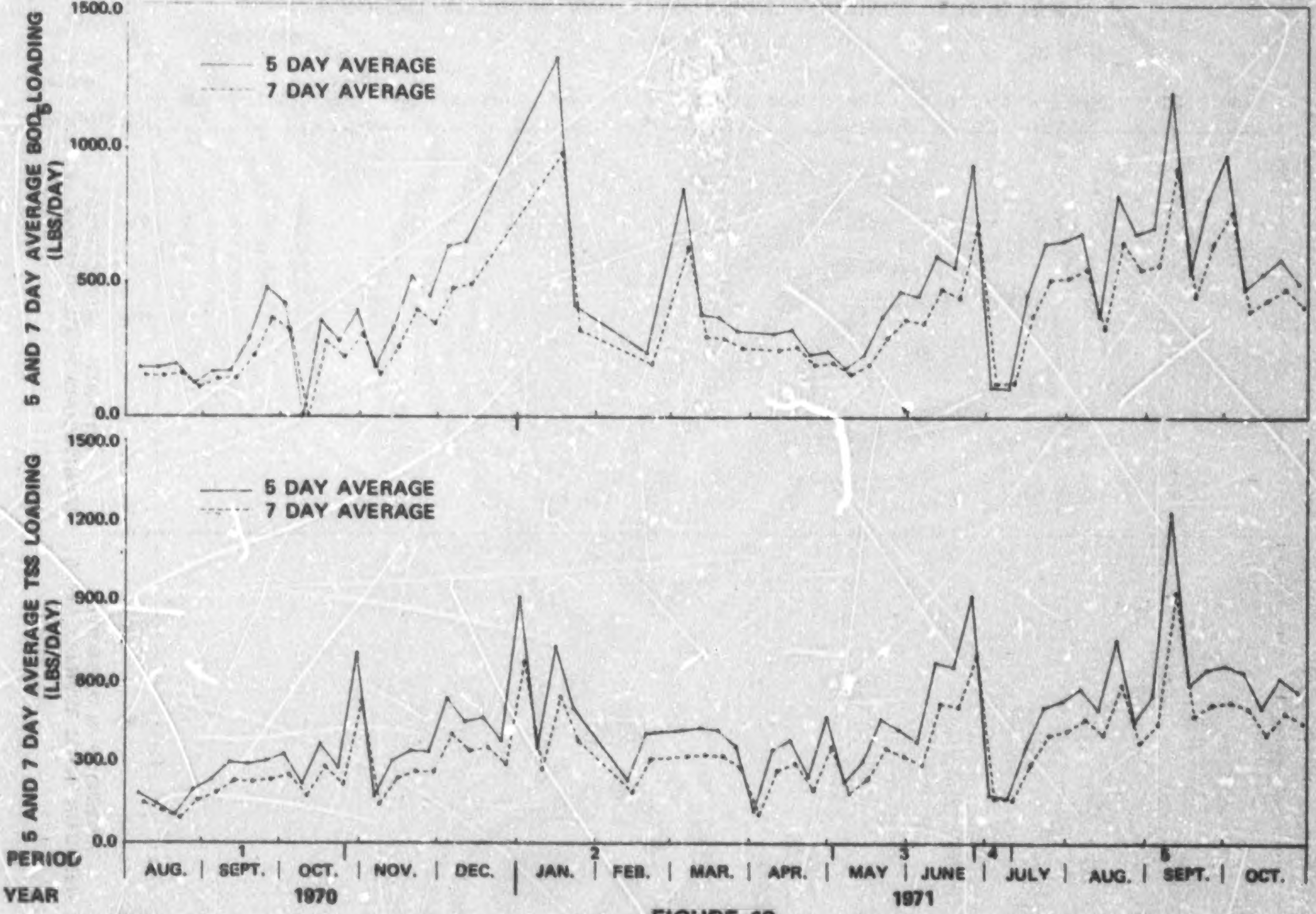


FIGURE 18

INFLUENT BOD AND TSS LOADINGS VERSUS TIME

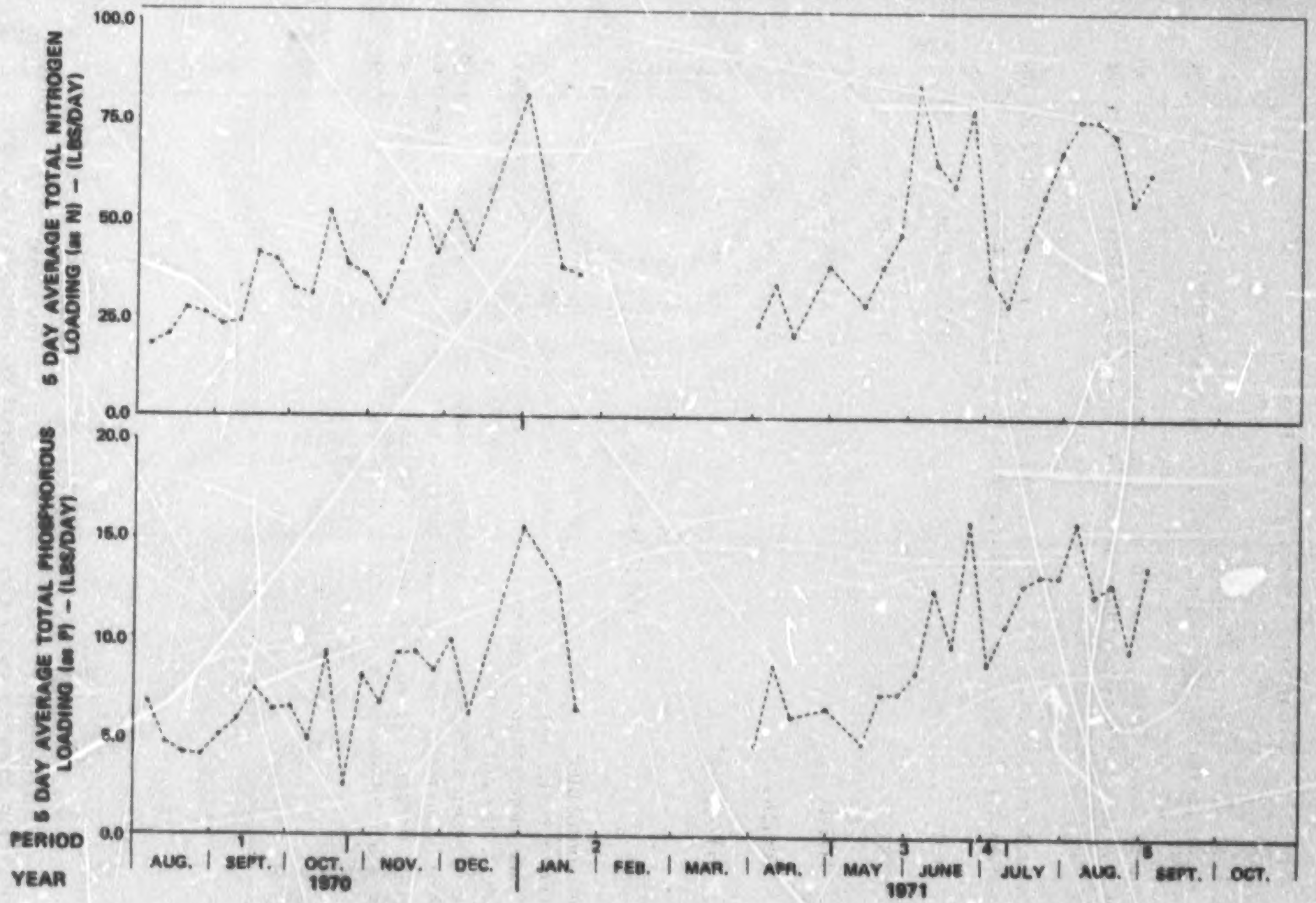


FIGURE 19
INFLUENT NITROGEN AND PHOSPHOROUS LOADINGS VERSUS TIME

DOMESTIC PER CAPITA LOADINGS

In August 1971, approximately 365 single-family dwellings were connected to the City sewerage system, contributing an average flow of 95,000 gpd to the treatment plant. The effect of infiltration on the total influent flow during the month was negligible due to the dry weather conditions of the preceding three months. The 1970 census indicated an average of 2.8 population per dwelling in the City of Tualatin. From these data the average per capita flow was calculated to be 93 gallons per day.

Using the average influent BOD, TSS, nitrogen, and phosphorus values determined for the domestic wastewater (Table 6), the following influent per capita loadings were calculated:

BOD	-	0.178 lb/cap/d
TSS	-	0.213 lb/cap/d
TOTAL - N	-	0.042 lb/cap/d
ORTHO - P	-	0.0086 lb/cap/d
TOTAL - P	-	0.0128 lb/cap/d

INFILTRATION

Infiltration contributed a significant flow to the treatment plant, during the prolonged wet weather of winter and spring. An estimate of the average daily infiltration flow was made for each month from November 1970 through May 1971. The infiltration flow was assumed to be the difference between the total plant flow and the combined industrial and domestic wastewater flows. The industrial wastewater contribution was measured directly. The domestic flow was estimated by applying the per capita flow of 93 gpcd to the estimated population connected to the plant each month. The average daily total plant flow and infiltration flow for each month and the total monthly rainfall are shown on Figure 20.

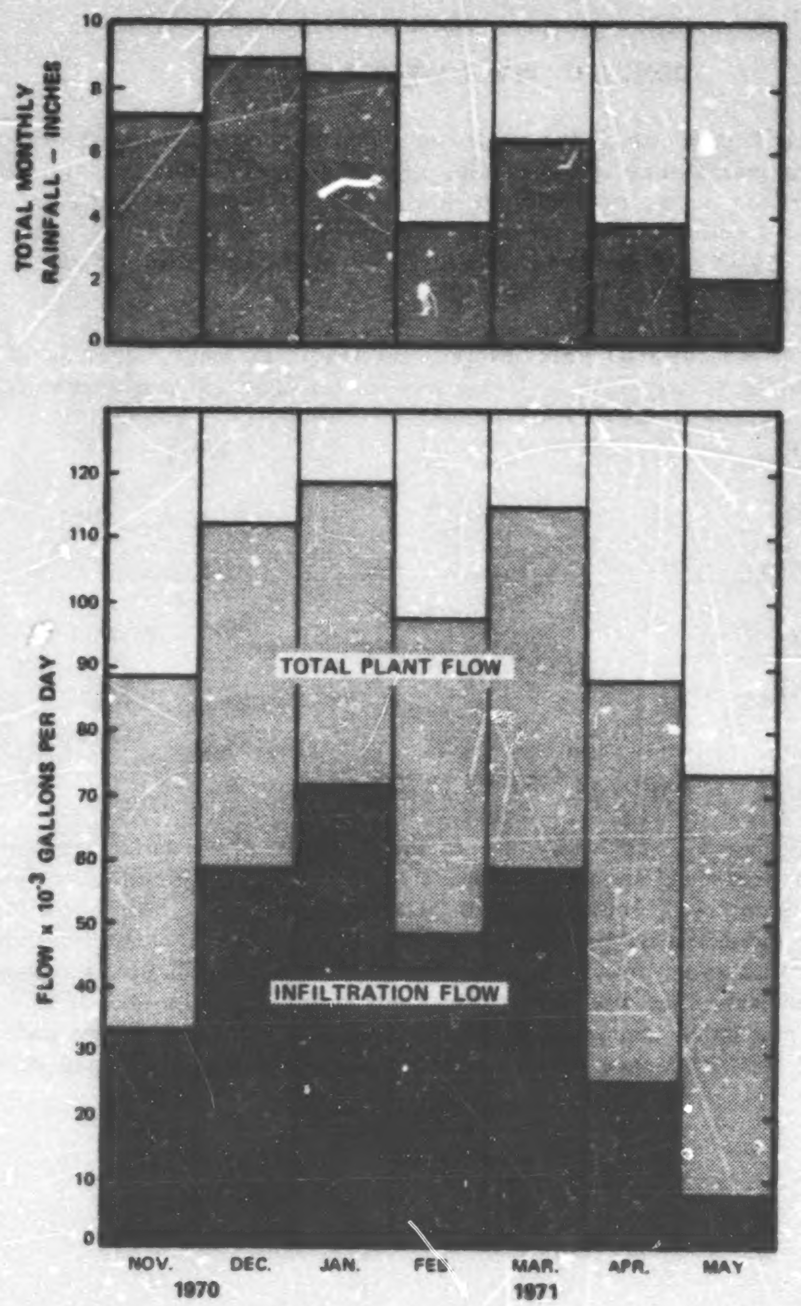


FIGURE 20

EFFECT OF INFILTRATION ON TOTAL PLANT FLOW

SECTION VII
TREATMENT PLANT PERFORMANCE

GENERAL PLANT PERFORMANCE

Although the plant was started up in April 1970, problems with the aeration-surge basin liner delayed routine testing and data analysis until August 1970. With the exception of a brief period in June and July when repairs were being made to the aeration-surge basin liner, the secondary portion of the treatment plant operated continuously during the demonstration project.

The tertiary system was operated on a routine basis from August 1970 through the end of November 1970, when freezing weather damaged several pieces of equipment. During December 1970 through March 1971, the tertiary unit was operated intermittently. With the onset of warmer weather in April the tertiary system was reactivated and run continuously for the duration of the project.

At the beginning of the project, both the hydraulic and organic loadings applied to the plant were considerably below the design capacity of the system.

During the course of the project, however, the organic loading frequently exceeded the aeration capability of the secondary system, resulting in a degradation in secondary effluent quality. The increased organic loading was primarily due to expansion of Hervin Company's production facilities and the inability of their pretreatment system to reduce the strength of the industrial wastewater to the level anticipated at the time the treatment plant was designed.

The continual addition of domestic connections to the Tualatin wastewater collection system, particularly during the months of June, July and August of 1971, added to the overload condition. This increase in domestic connections had been anticipated and allowed for in the design of the plant. The increased industrial waste load, however, used up most of the capacity allocated for the domestic flow.

The hydraulic capacity of the plant was never reached during the demonstration program. The maximum daily flow recorded was 0.256 mgd. During the peak flow month of January 1971, the average daily flow approached 45 percent of the design hydraulic capacity.

Two additional problems also affected secondary treatment performance. Excessive electrical current draw caused the aerators to kick off periodically, interrupting the oxygen supply to the aeration-surge basin. Although modifications were made to the aerators, this problem was never satisfactorily resolved during the demonstration program.

Entrapment of biological solids in the tube bundles of the secondary clarifier and subsequent septic conditions greatly increased the suspended solids content of the secondary effluent at various times in the first half of the program. Modifications made to the secondary tube clarifier in March 1971 substantially alleviated this problem.

Even though poor performance of the secondary system at certain times severely taxed the tertiary unit, a consistent, high quality final effluent could be maintained with the addition of chemicals.

SECONDARY TREATMENT SYSTEM

GENERAL

The following parameters were routinely monitored throughout the demonstration program to provide a basis for analysis of secondary treatment system performance:

Aeration-Surge Basin: Temperature, pH, alkalinity, dissolved oxygen (D.O.), mixed liquor suspended solids (MLSS), sludge volume index (SVI), and waste sludge.

Secondary Effluent[†]: BOD; TSS; kjeldahl, ammonia, nitrite, and nitrate nitrogen; ortho and total phosphorus, pH and alkalinity.

The secondary treatment system data was separated into the same five time periods for analysis and presentation as the plant influent data:

1. 3 August - 25 October 1970
2. 26 October 1970 - 2 May 1971
3. 3 May - 27 June 1971
4. 28 June - 11 July 1971
5. 12 July - 31 October 1971

Combined industrial and domestic wastewaters were treated during weekdays in periods 1, 2, 3, and 5. In period 4, Hervin Company's operations were suspended for annual employee vacations. Only domestic wastewater was treated in this two week period.

[†] Nitrogen and phosphorus determinations were not made in September and October 1971. Aeration-surge basin temperatures were not recorded in September 1971.

The secondary system performed reasonably well, producing an acceptable effluent for tertiary treatment most of the time during periods 1, 3, and 4. Solids accumulation problems in the secondary clarifier degraded the secondary effluent quality much of the time in period 2. The modifications made to the secondary tube clarifier considerably improved performance after March 1971.

In the final period, the organic loading frequently exceeded the aeration capacity of the plant. As a result, secondary treatment performance degenerated toward the end of the program.

AERATION-SURGE BASIN PERFORMANCE

The data obtained from the aeration-surge basin are summarized in Table 8. The average, maximum, minimum and standard deviations for the parameters measured are listed for each of the five time periods and for the entire demonstration project.

TEMPERATURE–The average weekly temperature of the aeration-surge basin varied from a maximum of 22.5 degrees C during the second week of August 1971 to a minimum of 8.1 degrees C in the second week of January 1971. The average temperature for the demonstration period was 15.5 degrees C. The daily aeration-surge basin temperature versus time is shown on Figure 21. The range of temperatures experienced by the secondary system did not appear to measurably affect treatment performance. The relatively long sludge ages and low F/M's tended to minimize the effects of temperature.

pH AND ALKALINITY–The pH and alkalinity were directly affected by the extent of nitrification in the aeration-surge basin. When D.O. levels were adequate to support full oxidation of the ammonia nitrogen fraction, the pH remained between 6.4 to 6.8 and the alkalinity between 40 to 80 mg/l. When nitrification was inhibited, the pH stabilized between 7.3 to 7.5 and the alkalinity remained above 200 mg/l. Continual monitoring of pH and alkalinity provided a reasonably good qualitative assessment of the extent of nitrification. The daily values of pH and alkalinity are plotted versus time on Figure 21.

Return of the tertiary backwash from the chemical sludge holding tank also affected the aeration-surge basin pH and alkalinity. During the first period (3 August-25 October 1970), low levels of alkalinity resulting from nitrification could not provide sufficient buffer capacity to offset the acidity of the backwash water, when alum was being fed to the tertiary system. Lime was added at the comminutor basin during September and October 1970 to control the aeration-surge basin pH.

To prevent pH depression in the aeration-surge basin during the spring and summer of 1971, the entire filter backwash volume was pumped directly to the chemical sludge storage lagoons. Algal activity maintained the pH of the lagoons above 7, effectively neutralizing the backwash water.

There was a tradeoff, however, to this mode of operation. The lagoon return supernatant was at times heavily laden with algae. A significant amount of algae appeared to pass

TABLE 8

AERATION–SURGE BASIN DATA SUMMARY

	TEMP. (DEG. C)	pH(1)	ALK. (MG/L)	D.O. (MG/L)	MLSS (MG/L)	SVI	WASTE SLUDGE(2) (LBS./DAY)	WASTE SLUDGE(2) (GPD)	SLUDGE AGE(3) (DAYS)	F/M(4)
PERIOD					3 AUGUST TO 25 OCTOBER 1970					
AVERAGE	17.2	6.90	43.	1.5	1210	80	263.	18,200	12.0	0.102
MAXIMUM	19.1	7.45	90.	3.9	1810	105	676.	41,710	31.1	0.198
MINIMUM	15.5	5.70	15.	0.2	920	46	95.	6,910	6.4	0.047
STD. DEV.	1.0	0.60	22.	1.0	246	15	187.	11,060	7.2	0.050
PERIOD					26 OCTOBER 1970 TO 2 MAY 1971					
AVERAGE	12.0	6.86	106.	1.6	1880	96	176.	7,060	12.3	0.144
MAXIMUM	16.1	7.85	223.	4.3	3130	134	711.	30,170	23.7	0.281
MINIMUM	8.1	6.05	34.	0.2	750	65	0	0	4.1	0.046
STD. DEV.	1.8	0.32	51.	1.1	706	18	179.	7,080	5.6	0.083
PERIOD					3 MAY TO 27 JUNE 1971					
AVERAGE	17.0	7.20	159.	0.8	2250	82	265.	9,770	17.6	0.098
MAXIMUM	20.0	7.40	230.	1.8	3250	93	815.	27,430	31.3	0.136
MINIMUM	14.2	6.80	112.	0.2	1805	71	0	0	8.1	0.038
STD. DEV.	1.9	0.16	39.	0.5	587	7	189.	8,900	8.0	0.033
PERIOD					28 JUNE TO 11 JULY 1971 (DOMESTIC WASTE ONLY)					
AVERAGE	16.0	6.80	85.	2.5	2060	99	260.	9,970	27.7	0.025
MAXIMUM	16.2	7.30	142.	2.9	2150	102	495.	18,680	31.1	0.026
MINIMUM	15.9	5.80	28.	2.0	1930	97	24.	1,260	24.4	0.024
STD. DEV.	0.2	0.60	80.	0.6	118	3	334.	12,320	4.7	0.001
PERIOD					12 JULY TO 31 OCTOBER 1971(5)					
AVERAGE	19.3	7.40	233.	0.4	2320	203	304.	16,510	12.0	0.136
MAXIMUM .	22.5	7.80	273.	2.1	2890	380	1154.	56,710	23.3	0.228
MINIMUM .	13.6	5.40	116.	0.1	1680	100	45.	1,710	4.6	0.080
STD. DEV.	2.8	0.28	50.	0.5	352	89	209.	13,620	4.8	0.041
PERIOD					TOTAL DEMONSTRATION PROJECT – 3 AUGUST 1970 – 31 OCTOBER 1971					
AVERAGE	15.5	7.40	118.	1.2	1910	116	260.	11,880	13.3	0.122
MAXIMUM	22.5	7.85	273.	4.3	3260	380	1154.	56,710	31.3	0.281
MINIMUM	8.1	5.40	15.	0.1	750	46	0	0	4.1	0.024
STD. DEV.	3.6	0.56	73.	1.0	656	68	216	11,150	6.6	0.063

(1) AVERAGE VALUE IS MEDIAN pH

(2) DAILY AVERAGE COMPUTED FROM WEEKLY AVERAGE.

(3) SLUDGE AGE DEFINED AS TOTAL SUSPENDED SOLIDS (LBS) IN THE AERATION–SURGE BASIN DIVIDED BY THE TOTAL SUSPENDED SOLIDS (LBS/DAY) LOST FROM THE SECONDARY TREATMENT SYSTEM. SLUDGE AGE COMPUTED ON A TWO WEEK MOVING AVERAGE.

(4) FOOD TO MICRO-ORGANISM RATIO (F/M) DEFINED AS TOTAL INFLUENT 5 DAY BOD (LBS/DAY) DIVIDED BY THE TOTAL SUSPENDED SOLIDS (LBS) IN THE AERATION–SURGE BASIN.

(5) TEMPERATURE RECORDED FOR JULY, AUGUST, AND OCTOBER ONLY.

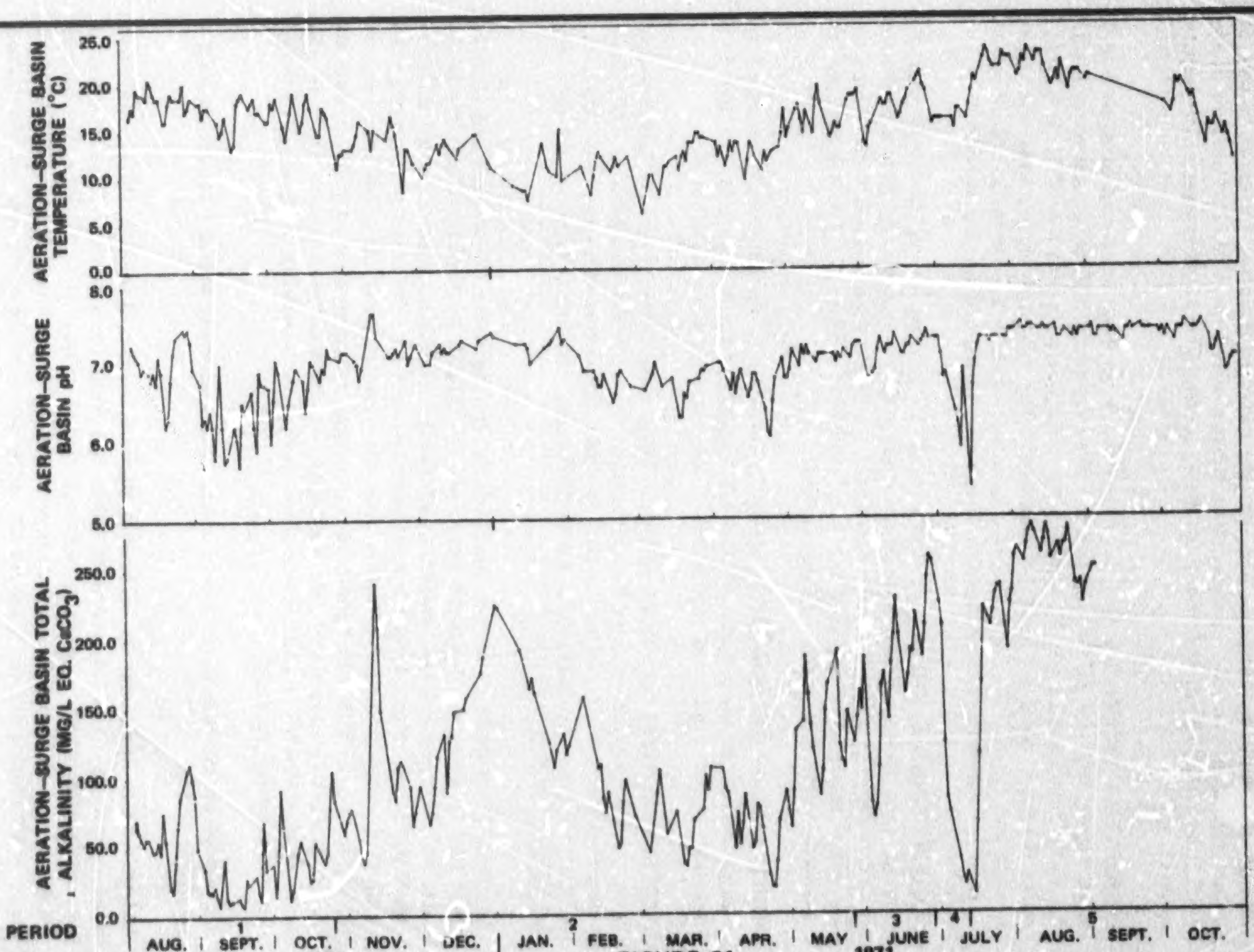


FIGURE 21
AERATION–SURGE BASIN TEMPERATURE, pH AND TOTAL ALKALINITY VERSUS TIME

through the aeration-surge basin and into the secondary effluent. The larger volume of lagoon return supernatant, resulting from putting all of the filter backwash in sludge holding lagoons, had the potential to increase the suspended solids loading to the tertiary system.

DISSOLVED OXYGEN–The D.O. in the aeration-surge basin fluctuated between 0 to 4 mg/l, varying with the influent BOD load. The daily values of D.O. are shown on Figure 22. From September to mid-November 1970 and January through May 1971, D.O. levels above 0.5 mg/l could generally be maintained. The increased organic load during the summer and fall of 1971 held the D.O. about 0.2 mg/l, except for the two week period in June and July when the industrial flow to the plant was temporarily suspended.

SLUDGE VOLUME INDEX–The SVI tended to increase during the demonstration program, ranging from a minimum weekly average of 46 in August 1970 to a maximum of 380 in October 1971. The average SVI for the demonstration period was about 116. The daily values of SVI versus time are shown on Figure 22. The increase in SVI appeared to correspond with a decrease in D.O.

MIXED LIQUOR SUSPENDED SOLIDS–The variation in the daily MLSS is shown on Figure 23. The weekly average of the MLSS ranged from a minimum of 750 mg/l to a maximum of 3,250 mg/l. The average MLSS for the program was 1,910 mg/l.

WASTE ACTIVATED SLUDGE–Biological solids were wasted from the aeration-surge basin through two suction lines located beneath the secondary tube clarifier. The mixing pattern created by the aerators tended to sweep the settled solids out from under the clarifier and into the basin. As a result, thickening was limited to about 1.2 times the MLSS concentration.

An attempt was made to thicken the solids by shutting down the aerators and allowing the MLSS to settle for 15 minutes before wasting. At SVI levels of 60 to 100, the MLSS could be thickened to between 6,000 to 9,000 mg/l. Thickening waste activated sludge in this manner had several limitations, however. After 15 to 30 minutes of wasting, the compacted sludge around the suction lines was withdrawn and the solids concentration rapidly decreased. With high organic loads frequently taxing aeration capacity, turning off the aerators only aggravated the difficulty of maintaining adequate D.O. in the aeration-surge basin. For these reasons, this method of thickening waste activated sludge was not generally practiced.

An average of 250 pounds per day of solids was intentionally wasted from the secondary treatment system.

SLUDGE AGE–The sludge age, defined as the total suspended solids in the aeration-surge basin divided by the total suspended solids lost from the secondary treatment system (secondary effluent and intentional wasting), ranged between 4 to 31 days. The average sludge age for the entire demonstration period was 13 days. Sludge age was calculated on a two week moving average. Data for the week in which the sludge age was determined was averaged with the data from the previous week. This method of calculation was

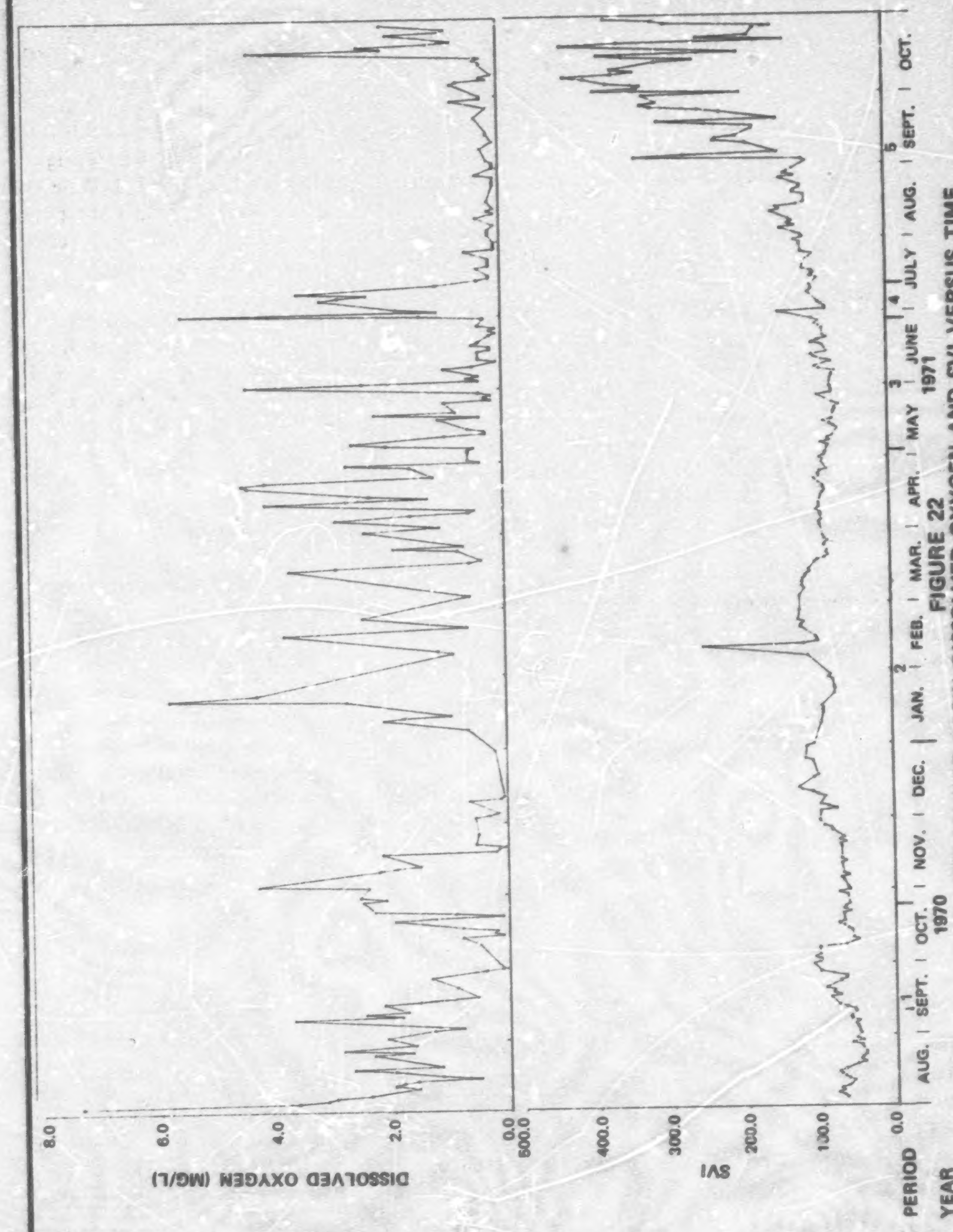


FIGURE 22
AERATION-SURGE BASIN DISSOLVED OXYGEN AND SVI VERSUS TIME

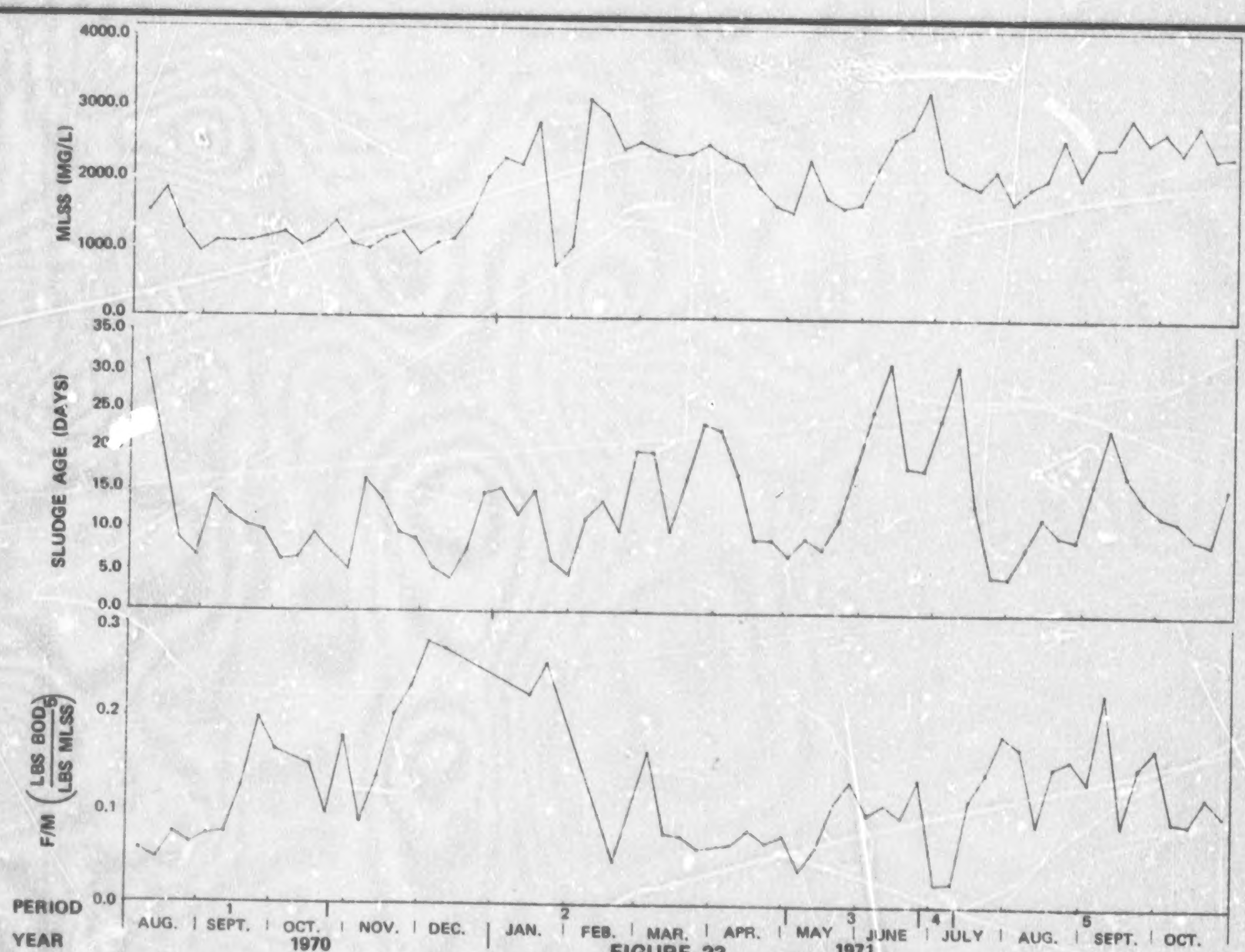


FIGURE 23
MLSS, SLUDGE AGE AND F/M VERSUS TIME

necessary to dampen the effects of infrequent activated sludge wasting. Sludge age versus time is shown on Figure 23.

F/M–The average weekly F/M varied between 0.024 to 0.28. The average F/M over the demonstration program was 0.122. F/M versus time is also shown on Figure 23.

DETENTION TIME–The hydraulic detention time in the aeration-surge basin averaged about 2.6 days, ranging from 7.1 days at the beginning of the program to about 1.2 days during peak flows in mid-January 1971.

SECONDARY EFFLUENT

The secondary effluent data are summarized in Table 9. The average, maximum, minimum and standard deviations of the parameters routinely measured are listed for each of the five time periods and for the total demonstration program.

BIOLOGICAL OXYGEN DEMAND–The BOD_5 removal in the secondary treatment system averaged 84 percent over the demonstration program. This poor BOD removal is attributed to several factors, including organic loadings in excess of the aeration capacity, aerator malfunction, and high solids carry over in the secondary effluent.

Nitrification was also suspected in many of the secondary and tertiary effluent BOD_5 tests, producing high readings. Simultaneous measurement of both the carbonaceous and nitrogeneous components in the BOD_5 analysis would considerably bias the test results. As a result, the percentage removals of the carbonaceous BOD fraction only were likely much higher than the test results indicated, particularly during May through October 1971.

Secondary effluent BOD_5 versus time is plotted on Figure 24. From August through October 1970 and for brief time intervals during other periods, total BOD_5 removals approached 92 to 95 percent.

To estimate actual soluble BOD_5 removal, a plot was made of the secondary effluent total BOD_5 versus suspended solids (Figure 25) for days when the D.O. was 1.0 mg/l or greater and nitrification had essentially gone to completion in the aeration-surge basin. From Figure 25 the average soluble BOD_5 concentration was determined to be about 3.0 mg/l, correlating well with tertiary effluent soluble BOD_5 measurements made during the same time periods. In terms of secondary effluent soluble BOD_5 removals exceeding 99 percent were possible, when adequate aeration was available.

SUSPENDED SOLIDS–The level of suspended solids in the secondary effluent varied widely, from a minimum of 20 mg/l to a maximum of 820 mg/l. The daily suspended solids content of the secondary effluent versus time is shown on Figure 24.

TABLE 9

SECONDARY EFFLUENT CHARACTERISTICS

	BOD$_5$		TSS		KJEL-N[1]	NH_3-N	ORG-N	NO_2-N	NO_3-N	TOTAL-N	ORTHO-P[2]	TOTAL-P	pH[3]	ALK[4]	DAILY FLOW
	(MG/L)	(LBS/DAY)	(MG/L)	(LBS/DAY)	(MG/L)	(MG/L)	(MG/L)	(MG/L)	(MG/L)	(MG/L)	(MG/L)	(MG/L)		(MG/L)	(MGD)
PERIOD	3 AUGUST – 25 OCTOBER 1970														
AVERAGE	28.	24.5	68.	53.6	10.5	1.5	9.0	0.41	16.6	27.4	6.90	7.8	7.05	39.	0.0730
MAXIMUM	111.	118.	477.	405.	36.0	6.4	33.6	3.64	32.0	47.7	12.0	15.7	7.40	104.	0.1876
MINIMUM	12.	1.7	21.	4.8	1.0	0.1	0.9	0.080	0.16	9.0	0.24	0.52	5.75	6.0	0.0114
STD. DEV.	20.	24.4	69.	69.7	6.3	1.5	5.7	0.63	8.7	9.2	3.60	3.8	0.46	78.	0.0499
PERIOD	26 OCTOBER 1970 – 2 MAY 1971														
AVERAGE	102.	110.	187.	184.	17.8	5.9	11.9	0.38	8.8	27.0	3.84	7.0	7.10	77.	0.1030
MAXIMUM	400.	676.	820.	1,260	46.3	24.1	41.3	2.05	22.1	61.7	7.92	13.0	7.86	208.	0.2240
MINIMUM	17.	7.7	25.	11.7	1.8	0.2	1.5	0.002	0.01	12.5	0.90	1.32	4.50	8.0	0.0302
STD. DEV.	101.	137.	220.	248.	13.9	7.3	9.8	0.49	6.7	12.5	1.98	2.9	0.42	44.	0.0386
PERIOD	3 MAY – 27 JUNE 1971														
AVERAGE	74.	57.3	56.	43.0	22.0	10.7	11.3	0.23	2.1	24.4	4.45	6.0	7.90	130.	0.0849
MAXIMUM	151.	143.	141.	134.	36.1	20.5	25.8	0.98	19.0	35.3	7.80	9.7	7.60	193.	0.1658
MINIMUM	22.	16.0	26.	10.0	6.4	0.1	5.3	0.005	0.005	12.3	1.79	2.05	6.95	50.	0.0256
STD. DEV.	36.	38.1	28.	29.8	9.3	6.7	5.8	0.30	5.1	7.4	2.07	1.9	0.17	37.	0.0361
PERIOD	28 JUNE – 11 JULY 1971 (DOMESTIC WASTE ONLY)														
AVERAGE	65.	40.8	81.	31.5	14.9	6.6	8.3	8.4	14.7	38.0	6.62	8.0	6.9	69.	0.0746
MAXIMUM	104.	86.4	404.	92.0	44.7	24.1	30.0	25.2	33.6	44.8	9.40	10.5	7.60	181.	0.1200
MINIMUM	29.	10.6	22.	9.9	2.7	0.1	2.5	0.069	0.03	31.3	3.60	4.70	6.20	9.0	0.0270
STD. DEV.	31.	32.3	121.	24.2	16.0	9.4	9.0	10.2	15.5	4.2	2.32	2.2	0.58	71.	0.0300
PERIOD	12 JULY – 31 OCTOBER 1971[5]														
AVERAGE	79.	82.5	108.	106.	36.2	22.0	14.2	0.20	3.3	39.6	6.52	8.3	7.40	192.	0.1101
MAXIMUM	178.	202.	586.	586.	60.8	34.0	45.2	3.80	38.2	60.9	12.6	15.0	7.56	257.	0.2561
MINIMUM	27.	26.5	20.	18.7	2.9	0.1	0.9	0.006	0.003	11.7	2.10	3.91	6.80	44.	0.0429
STD. DEV.	40.	45.8	96.	94.	15.5	8.4	9.7	0.86	10.0	11.5	3.01	3.4	0.14	46.	0.0268
PERIOD	TOTAL DEMONSTRATION PROJECT – 3 AUGUST 1970 – 31 OCTOBER 1971														
AVERAGE	73.	71.9	116.	107.	21.5	10.3	11.2	0.93	9.1	31.4	5.88	7.6	7.5	111.	0.0973
MAXIMUM	400.	676.	820.	1,260.	60.8	34.0	45.2	25.2	38.2	61.7	12.6	15.7	7.86	257.	0.2561
MINIMUM	12.	1.7	20.	4.8	1.0	0.1	0.9	0.002	0.003	9.0	0.24	0.52	4.5	6.0	0.0114
STD. DEV.	96.	85.0	148.	182.	15.5	10.9	7.7	3.5	10.7	11.3	3.15	3.2	0.55	73.	0.0391

(1) NITROGEN REPORTED AS N (M.W. 14.01)

(2) PHOSPHORUS REPORTED AS P (M.W. 30.98)

(3) AVERAGE VALUE IS MEDIAN pH

(4) ALKALINITY REPORTED AS EQUIVALENT $CaCO_3$ (M.W. 100.09)

(5) N AND P DETERMINED FOR JULY AND AUGUST ONLY.

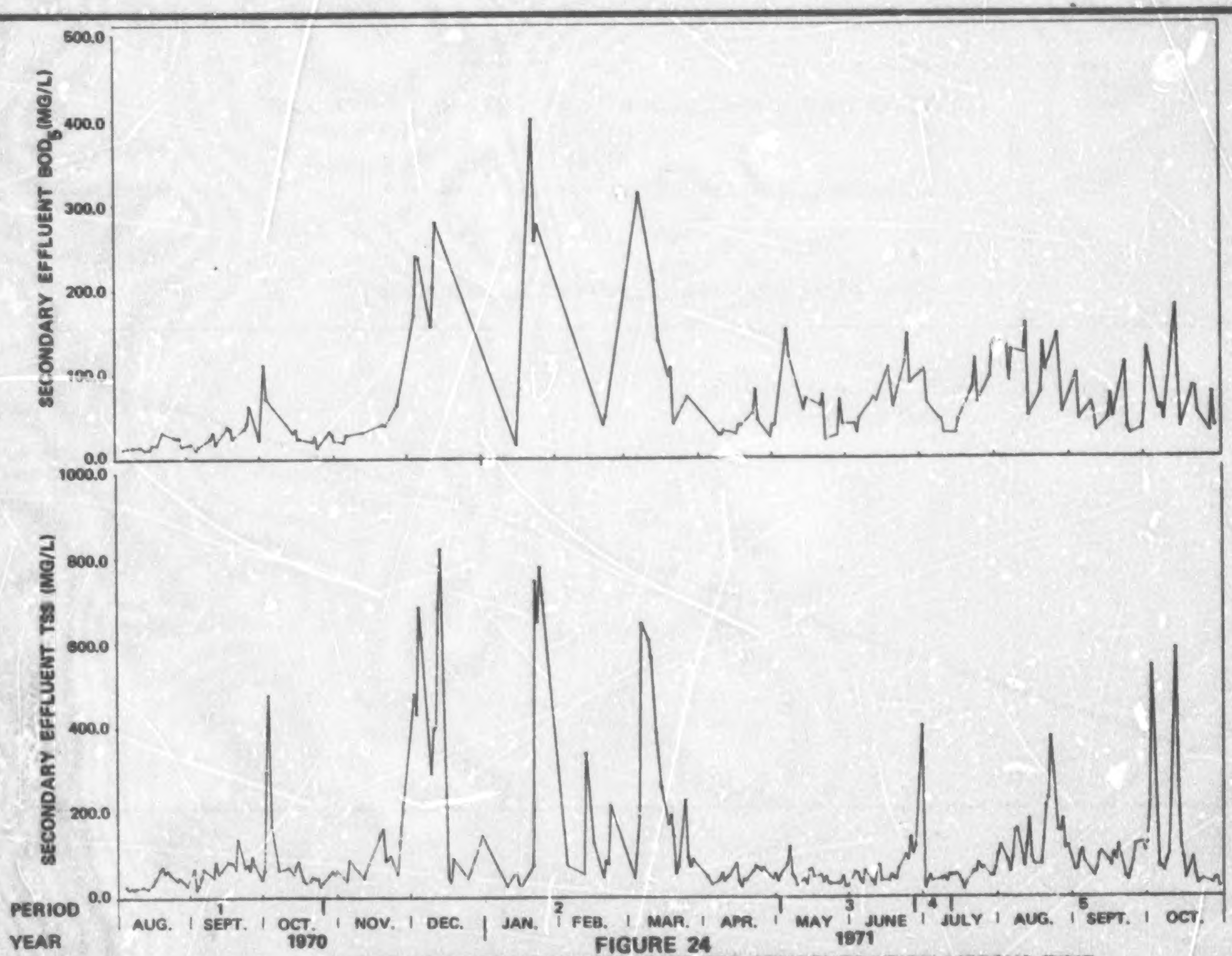


FIGURE 24
SECONDARY EFFLUENT BOD AND TSS CONCENTRATION VERSUS TIME

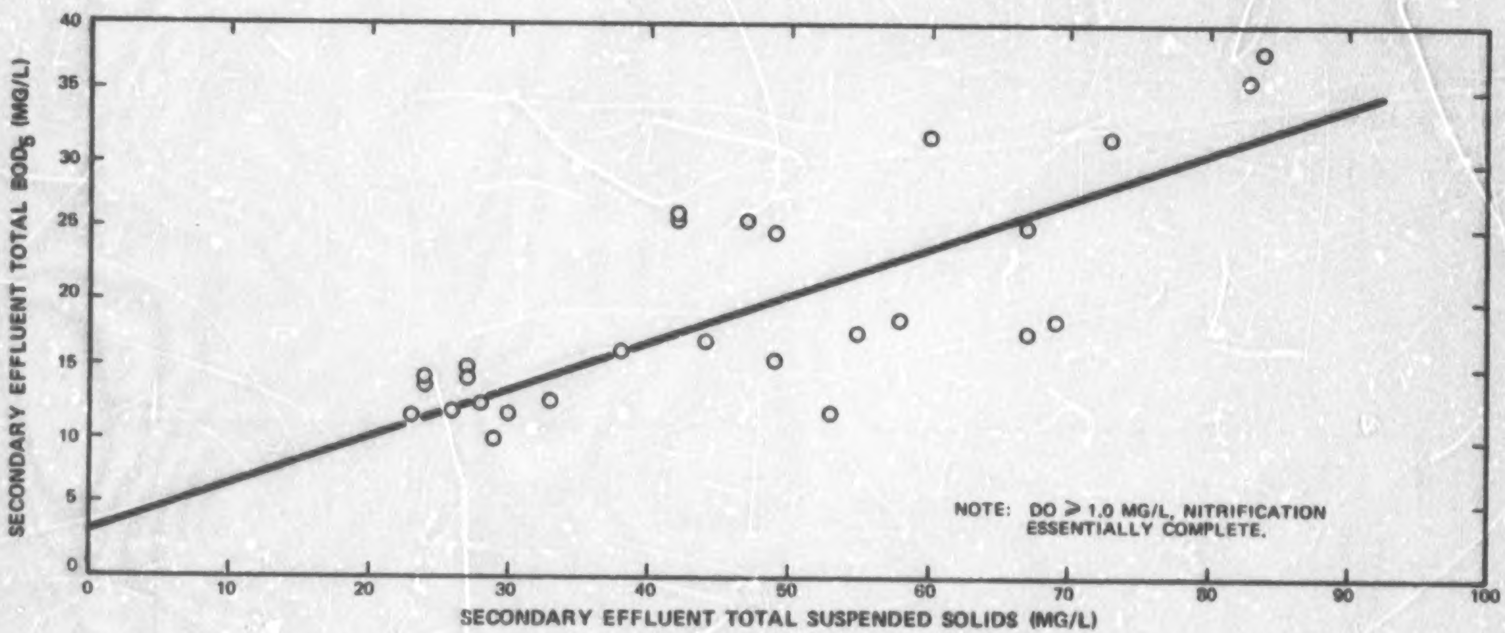


FIGURE 25

SECONDARY EFFLUENT TOTAL BOD VERSUS SUSPENDED SOLIDS

With the exception of the first week in October 1970, the secondary effluent suspended solids level remained consistently below 85-90 mg/l and averaged 60 mg/l from August through mid-November 1970. From mid-November until the air sparging modifications to the secondary clarifier were completed in March, large quantities of solids were periodically carried over, presumably as a result of solids accumulation and gassification in the tube modules.

Daily sparging for 15 minutes prevented solids build-up in the tubes and considerably improved the effluent quality. Except for a two week period toward the end of June 1971, the secondary effluent suspended solids concentration averaged 48.5 mg/l until the last week of July 1971. From this point on through the end of the program, the secondary effluent quality degenerated primarily as a result of the inability to maintain adequate D.O. levels in the aeration-surge basin.

The suspended solids removal in the secondary treatment system averaged 75 percent over the demonstration program. For the period following modification of the secondary clarifier, the suspended solids removal averaged 84 percent.

The possible effects of MLSS concentration, temperature, surface overflow rate, pH and extent of nitrification on the secondary tube clarifier performance and effluent suspended solids concentration are discussed in Section VIII.

NITROGEN–The total, ammonium and nitrate concentrations measured in the secondary effluent are shown versus time on Figure 26. The long sludge ages, at which the secondary treatment system was operated, presented ideal conditions for nitrification. So long as sufficient D.O. was available, nitrification approached completion. The extent to which D.O. controlled nitrification can be seen in comparing the D.O. levels on Figure 22 with the ammonium and nitrate concentrations on Figure 26. At D.O. levels above 1.0 mg/l the ammonium concentration generally remained below 1.0 mg/l. Reducing the D.O. to around 0.2 to 0.4 mg/l effectively inhibited nitrification, limiting the nitrate concentration to less than 0.5 mg/l.

The total nitrogen content of the secondary effluent ranged between 9.0 and 61.7 mg/l and averaged 31.4 mg/l, representing a 42 percent reduction in influent total nitrogen. The higher total nitrogen levels were primarily due to large suspended solids carryover in the secondary effluent.

The soluble nitrogen content – ammonium, nitrite and nitrate – varied between 2.6 and 40.4 mg/l and averaged 18.5 mg/l. Filtering the secondary effluent to remove suspended solids could potentially have increased the total nitrogen removal to 66 percent.

PHOSPHORUS–The ortho and total phosphate concentrations of the secondary effluent versus time are also shown on Figure 26. The secondary effluent total phosphate concentration ranged from 0.52 to 15.7 mg/l and averaged 7.6 mg/l. This represents an average reduction of 32 percent of the influent total phosphorous load.

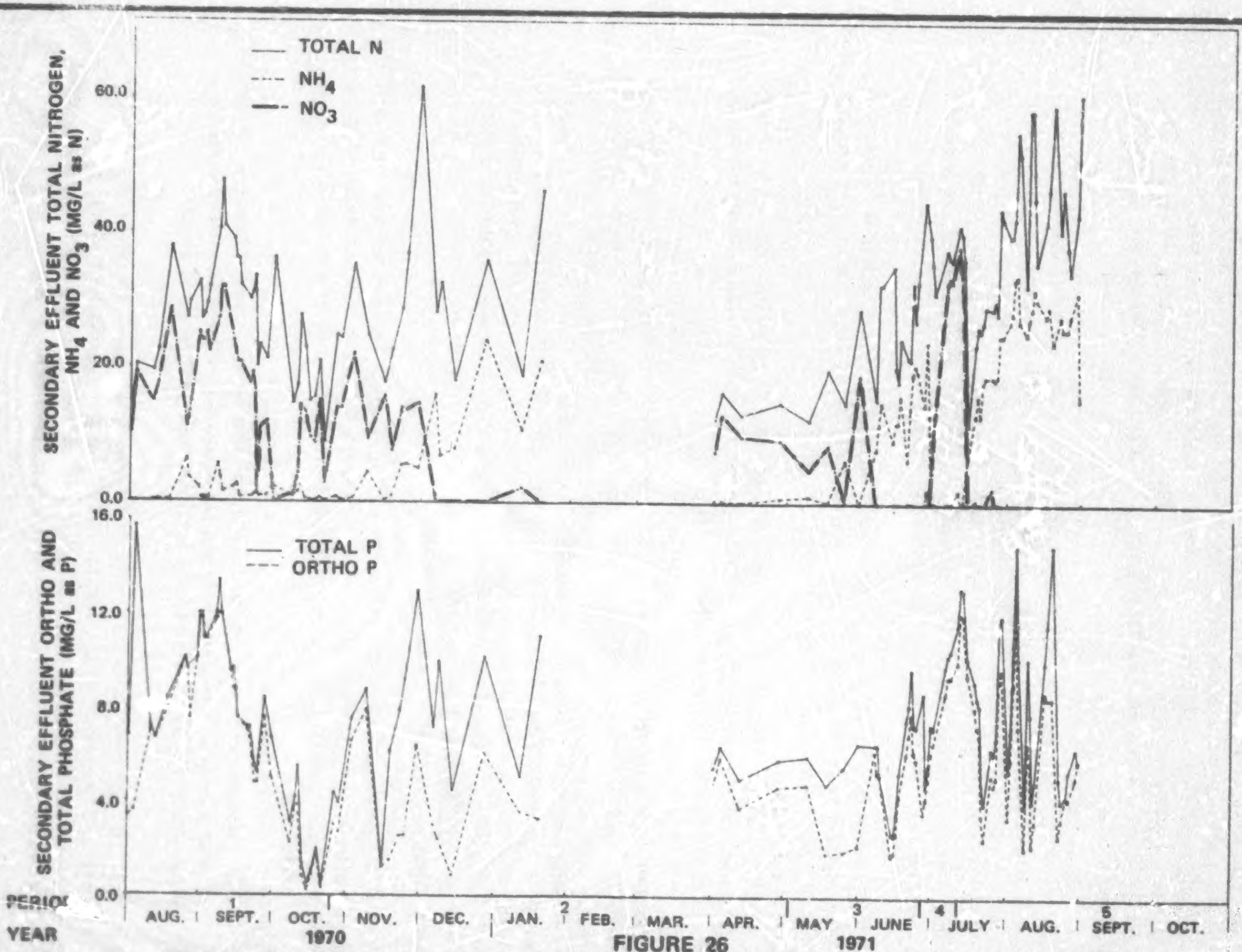


FIGURE 26

SECONDARY EFFLUENT NITROGEN AND PHOSPHOROUS CONCENTRATIONS VERSUS TIME

The orthophosphate varied from a minimum of 0.24 mg/l to a maximum of 12.6 mg/l and averaged 5.88 mg/l.

The difference in the ortho and total phosphate concentrations is believed to be due to the high level of suspended solids in the secondary effluent. The hydraulic detention times in the aeration-surge basin were such that polyphosphates would likely have been hydrolyzed to orthophosphate. The negligible difference between the ortho and total phosphate in the tertiary effluent, when virtually all of the suspended solids were removed, tends to support this conclusion.

The total influent phosphate and secondary effluent orthophosphate loadings were averaged on a weekly basis and compared to evaluate phosphate removal in the aeration-surge basin, assuming complete separation of secondary effluent suspended solids. The phosphate removal varied considerably from 0 to 87 percent and averaged 48 percent. The possible effects of pH and alkalinity on phosphate removal in the secondary treatment system are discussed in Section VIII.

VELOCITY PROFILES

Velocity measurements were made at 21 locations in the northeast quadrant of the aeration-surge basin at depths of 1, 6 and 9 feet. The velocity profiles shown on Figures 27, 28, and 29 were constructed from these data. Since the basin was symmetrical, the velocity profile of the northeast quadrant was assumed to be typical of the other three quadrants.

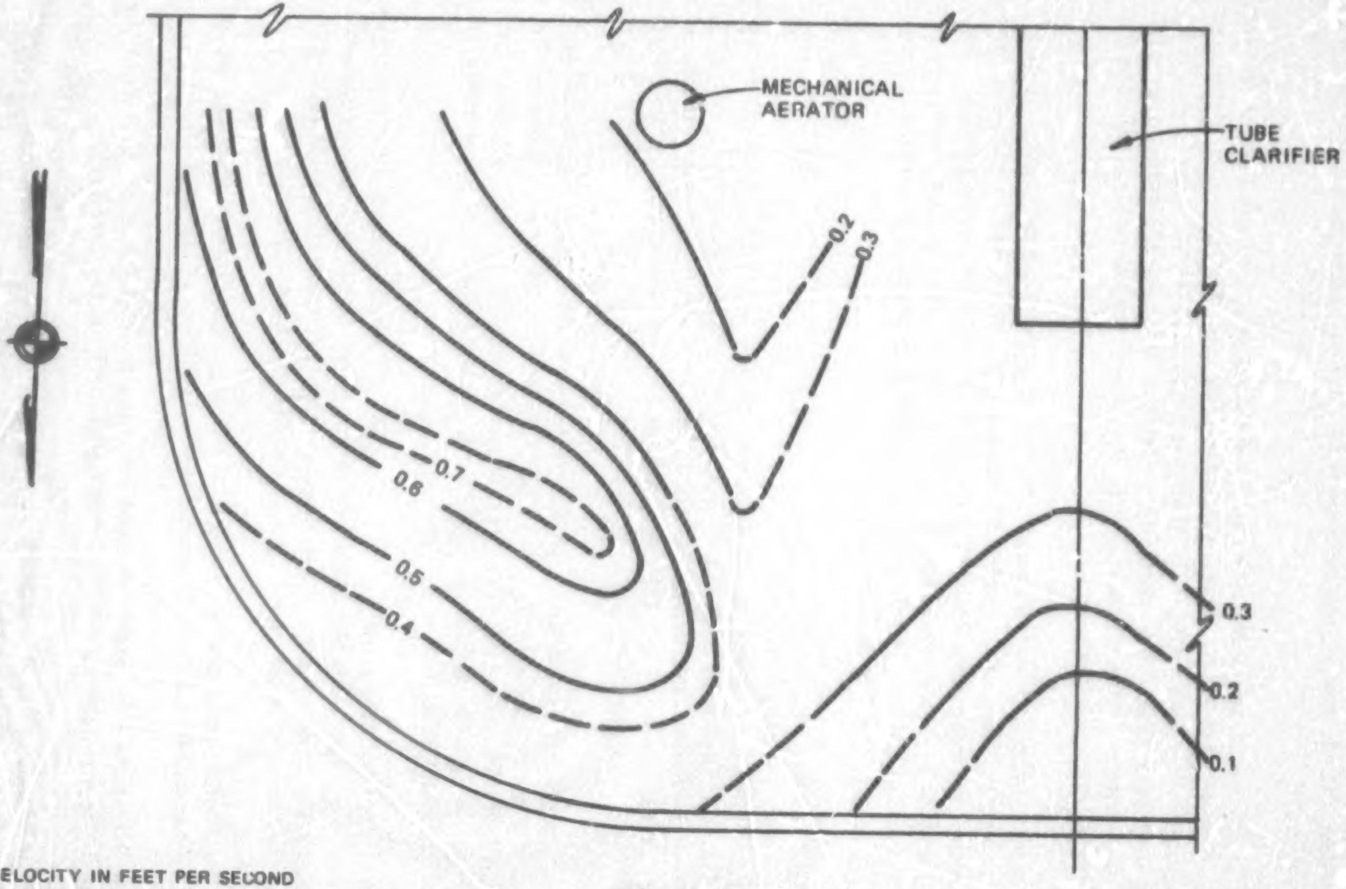
Velocity measurements were made with a Hydro Products Model 451 current meter. Only the horizontal component of velocity could be determined with this instrument. The resultant velocity of both the horizontal and vertical components may well have been higher than those indicated on the profiles. Both aerators were operating when the velocity measurements were made.

At a depth of 1 foot below the surface, the minimum horizontal velocity of 0.1 feet per second occurred at a point along the secondary effluent withdrawal pipe approximately 5 feet inboard from the aeration-surge basin ringwall. The maximum horizontal velocity of 0.7 feet per second was measured about 20 feet radially out from the aerator.

The horizontal velocities measured at a depth of 6 feet ranged from 0.7 feet per second at basin liner, decreasing to 0.3 feet per second within 5-10 feet of the aerator.

At a depth of 9 feet the horizontal velocities increased from 0.3 feet per second at the liner to about 0.6 feet per second within 4 feet horizontally of the aerator.

The velocity data indicate that adequate mixing was provided throughout the basin. Solids deposition would be expected in the low velocity areas between the ends of the secondary clarifier and the basin ringwall.

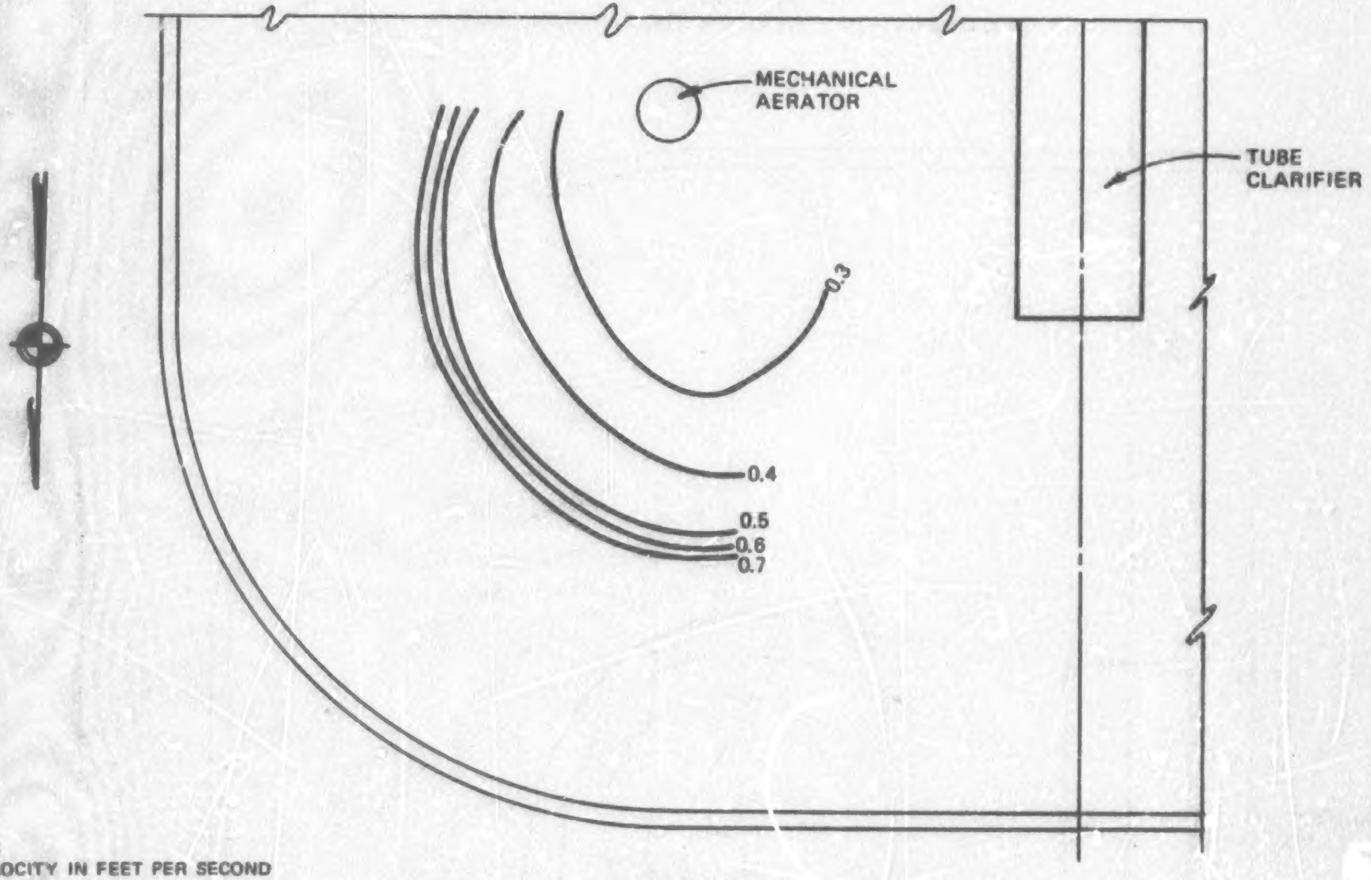

MECHANICAL AERATOR
TUBE CLARIFIER
0.2
0.3
0.7
0.6
0.5
0.4
0.3
0.2
0.1

NOTES:

1. VELOCITY IN FEET PER SECOND
2. DASHED LINE INDICATES CONTOUR BASED ON INTERPOLATION.

FIGURE 27

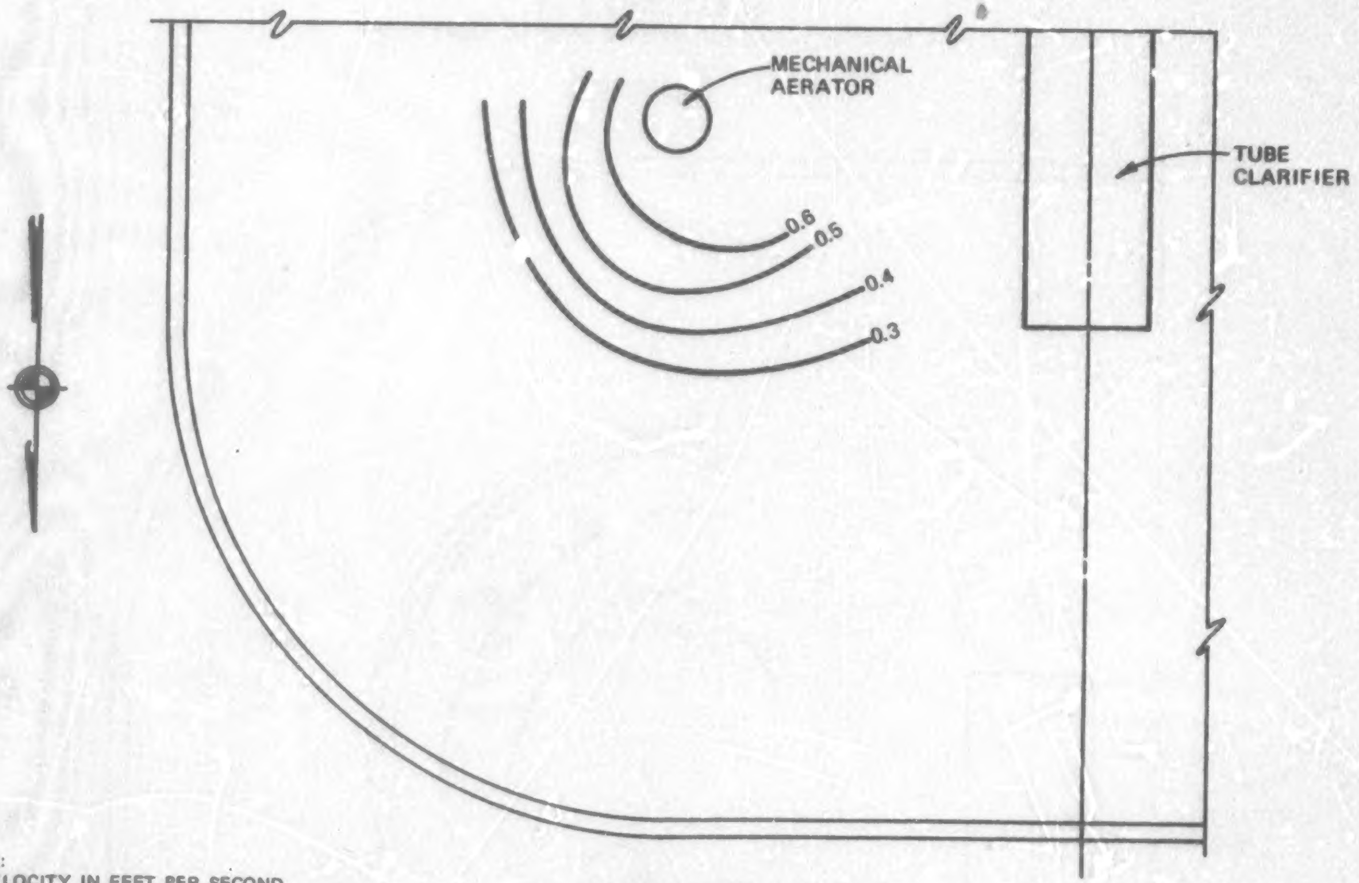
AERATION–SURGE BASIN VELOCITY PROFILE AT 1 FOOT DEPTH


MECHANICAL
AERATOR
TUBE
CLARIFIER
0.3
0.4
0.5
0.6
0.7

NOTE:
VELOCITY IN FEET PER SECOND

FIGURE 28

AERATION–SURGE BASIN VELOCITY PROFILE
AT 6 FOOT DEPTH


MECHANICAL
AERATOR
TUBE
CLARIFIER
0.6
0.5
0.4
0.3

NOTE:
VELOCITY IN FEET PER SECOND

FIGURE 29

AERATION–SURGE BASIN VELOCITY PROFILE
AT 9 FOOT DEPTH

SOLIDS ACCUMULATION

A high ground water table prevented dewatering of the aeration-surge basin to determine the solids accumulation. An attempt was made to profile the material deposited on the bottom of the basin by taking soundings. A weighted flat plate was suspended by a calibrated rope and lowered through a pulley system until the plate rested on the basin floor. Using this method the northeast quadrant was sounded at 21 locations.

The area of deposition is shown on Figure 30.

The solids accumulation was found to be minimal. The depth of the material varied from about 3 inches at a distance of 7 feet measured radially from the aerator to about 12 inches at the north end of the secondary clarifier. The average depth was estimated to be 6 inches. The location of the deposit is consistent with velocity profiles determined for the aeration-surge basin.

From these measurements and the total flow recorded for the 17-1/2 months the basin was in service, the solids accumulation rate was conservatively estimated to be about 0.32 cubic yards per million gallons. This value is within the range of grit loading experienced by other sewage treatment plants in the area.

TERTIARY TREATMENT SYSTEM

GENERAL

Alum coagulation supplemented with anionic polyelectrolyte, tube sedimentation and mixed media filtration was demonstrated to be an effective process for removing suspended solids and phosphate from the effluent of an extended aeration activated sludge system. The physical and chemical characteristics of the secondary effluent varied widely. This variability was attributed to the wide range of oxidative and sludge age conditions in which the secondary treatment system operated and certain performance limitations of the secondary tube clarifier.

The following tertiary effluent parameters were monitored on a regular basis throughout the demonstration program:

Temperature

pH

Total alkalinity

D.O.

TSS

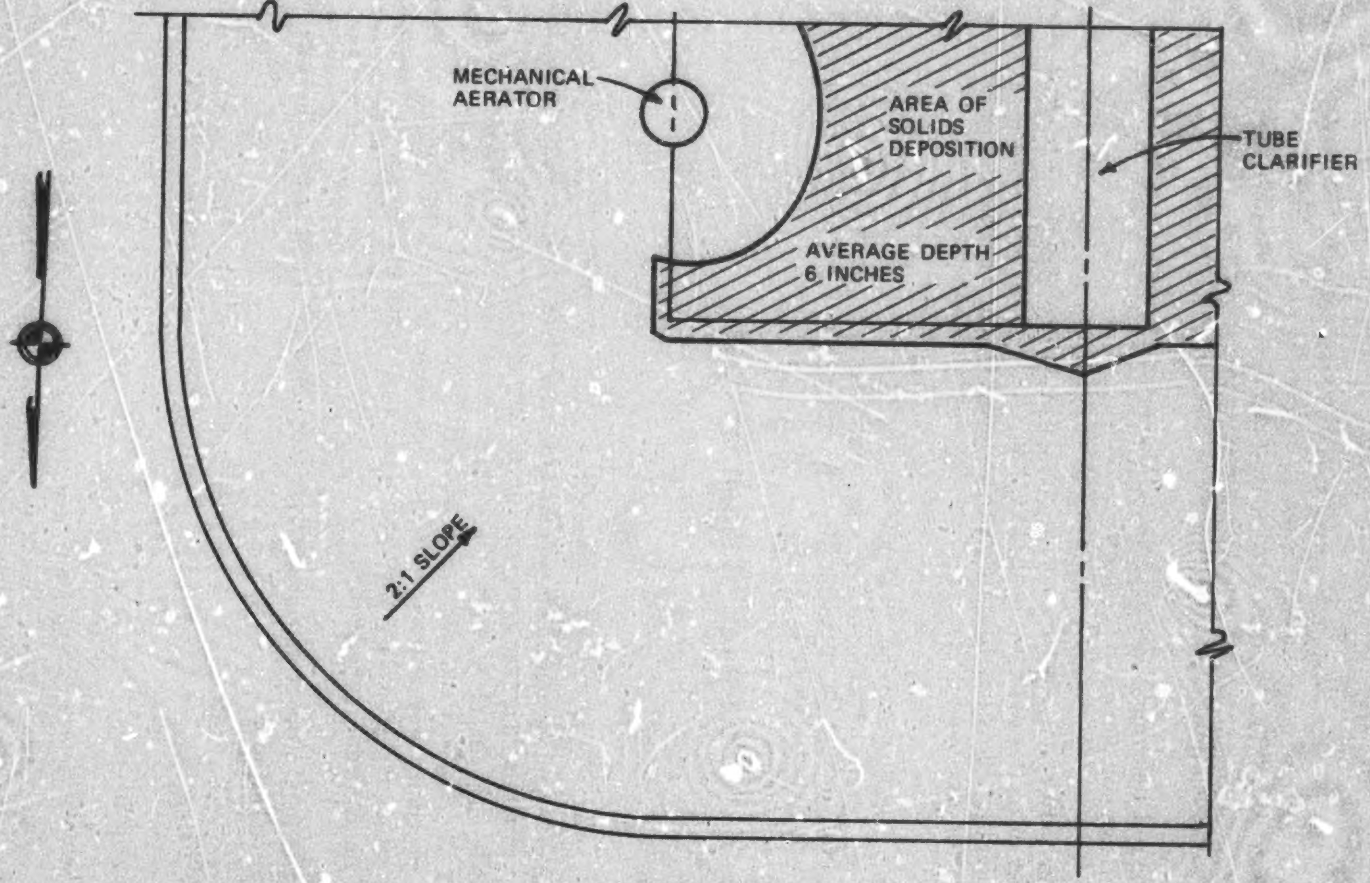

MECHANICAL AERATOR
AREA OF SOLIDS DEPOSITION
TUBE CLARIFIER
AVERAGE DEPTH 6 INCHES
2:1 SLOPE

FIGURE 30

AERATION–SURGE BASIN SOLIDS ACCUMULATION

Total BOD_5

Dissolved BOD_5

Ortho and total phosphate

Kjeldahl, ammonia, nitrite and nitrate nitrogen

Turbidity, filter run time, and filter head loss were also monitored on a regular basis from March 1971 through completion of the project.

Tertiary system performance is summarized in Table 10. The average, maximum and minimum values of the tertiary effluent quality parameters, throughput volume and solids removal per filter cycle are listed for various alum feed conditions. Because of the limited amount of data for each alum dosage range, standard deviations were not computed.

JAR TESTS

Jar tests were performed at the beginning and repeated several times during the course of the project to establish the alum dosages necessary for coagulation and phosphate removal. The alum dosage required to effect coagulation varied between 50 to 140 mg/l depending on the alkalinity and pH of the secondary effluent. The orthophosphate concentration could generally be reduced to less than 1 mg/l with alum dosages of 80 to 120 mg/l. The degree of phosphate removal appeared to be strongly affected by pH, with optimum removals occurring in a pH range of 5 to 6.

A series of jar tests was also run to evaluate the effect of polyelectrolyte on coagulation and floc formation. Initially, several anionic, nonionic and cationic polyelectrolytes were screened to determine the type best suited to complement the coagulation process. Anionic polymers enhanced the formation of a large cohesive floc. Nonionic polymers had little or no effect on floc formation and cation ionic polymers appeared to act as a dispersant.

Anionic polymers of several manufacturers were then evaluated to select a polyelectrolyte to be used during the project. The polymers tested included Dow Chemical Company Purifloc A21 and A23; American Cyanamide Company Magnifloc 835A, 836A and 837A; and Rohm and Haas Company Primafloc A10. Of these, A23, 836A and A10 produced the best results in aiding floc formation at dosages of 1.0 to 2.0 mg/l. American Cyanamide 836A was selected as the polyelectrolyte for use in the tertiary system, because of its availability from a local distributor near the treatment plant.

CHEMICAL FEED RATES

During August 1970 and for several brief intervals during the project, the tertiary system was operated without chemicals to establish a performance base line for the unit. On 31 August 1970, alum was introduced into the tertiary influent and on 9 October polyelectrolyte was added to the flocculator.

TABLE 10

TERTIARY PERFORMANCE SUMMARY

	BOD_5			FLOW	TSS	ORTHO–P (as P)		TOTAL P (as P)		TURBIDITY
	SOL. (MG/L)	TOT. (MG/L)	TSS (MG/L)	VOLUME (GAL./RUN)	REMOVED (LBS./RUN)	(MG/L)	% REM.	MG/L	% REM.	(JTU)
					NO CHEMICALS					
AVERAGE	5.0	12.1	12.3	148,700	106.	6.07	17.3	6.28	32.7	3.2
MAXIMUM	8.6	19.2	28.0	478,000	237.	8.30	28.6	8.90	38.6	3.5
MINIMUM	2.8	7.0	2.3	38,000	23.2	3.40	2.4	3.45	23.3	2.7
					ALUM 63 MG/L					
AVERAGE	–	–	2.7	33,600	14.8	5.28	45.9	5.22	53.1	1.8
MAXIMUM	–	–	4.5	44,700	19.7	6.56	57.4	6.35	61.0	2.0
MINIMUM	–	–	.9	22,200	9.8	4.00	34.4	4.10	45.2	1.3
					ALUM 60–63 MG/L					
AVERAGE	–	8.3	1.4	50,100	22.7	1.10	79.2	1.12	81.9	1.3
MAXIMUM	–	11.8	2.8	78,100	41.6	1.40	85.6	1.40	85.8	–
MINIMUM	–	4.8	.6	23,100	4.6	.80	72.9	.83	78.0	–
					ALUM 94-96 MG/L					
AVERAGE	6.2	8.8	6.8	46,000	34.5	1.96	73.2	2.05	75.9	1.6
MAXIMUM	13.9	17.3	22.0	57,500	40.7	6.05	93.3	6.0	96.1	2.0
MINIMUM	2.6	3.5	0.7	34,200	22.5	0.19	45.5	0.24	54.5	1.3
					ALUM 105–108 MG/L					
AVERAGE	4.1	7.8	1.6	42,700	35.4	1.11	84.2	1.22	86.2	1.1
MAXIMUM	5.2	30.0	10.4	67,200	123.	1.99	93.0	2.20	94.4	1.8
MINIMUM	3.0	0.5	0.2	25,200	13.7	.23	75.4	0.23	76.1	0.8
					ALUM 126 MG/L					
AVERAGE	4.0	4.4	3.2	41,100	22.5	0.68	89.8	0.69	91.4	1.0
MAXIMUM	8.4	8.9	13.1	60,800	99.0	2.20	97.9	2.20	98.4	2.1
MINIMUM	1.9	1.8	0.1	20,900	7.3	0.09	68.1	0.11	73.2	0.4
					ALUM 135 MG/L					
AVERAGE	2.2	2.6	0.7	48,300	40.5	0.16	96.2	0.16	96.4	1.2
MAXIMUM	3.5	3.5	2.0	58,600	58.6	0.25	98.3	0.25	98.6	–
MINIMUM	1.0	1.7	0.0	41,800	22.6	0.07	94.2	0.07	94.2	–
					ALUM 165 MG/L					
AVERAGE	4.3	5.3	2.1	36,500	16.1	0.65	86.7	0.73	88.0	.6
MAXIMUM	9.2	9.8	15.0	57,000	46.2	3.57	99.6	3.84	99.0	1.8
MINIMUM	1.6	2.6	0.0	25,600	4.9	0.01	8.3	0.03	8.8	.35
					ALUM 182–184 MG/L					
AVERAGE	4.4	5.5	1.3	33,200	11.8	0.12	96.8	0.16	96.9	0.7
MAXIMUM	10.8	11.8	3.2	40,400	17.5	0.25	98.9	0.27	97.9	0.9
MINIMUM	2.2	3.4	0.1	24,700	7.8	0.02	95.5	0.04	94.8	0.5
					ALUM 211–217 MG/L					
AVERAGE	8.5	7.4	1.4	31,600	24.1	0.23	95.1	0.26	95.1	0.65
MAXIMUM	17.6	19.1	4.0	41,800	38.2	0.65	98.4	0.66	97.8	0.85
MINIMUM	0.8	1.3	0.1	14,200	9.1	0.01	90.6	0.02	92.0	0.55
					ALUM 238–242 MG/L					
AVERAGE	6.1	9.4	2.0	34,300	33.2	0.34	93.6	0.39	94.8	1.0
MAXIMUM	13.0	19.0	9.3	50,900	81.5	1.87	98.1	1.90	97.9	2.4
MINIMUM	1.3	2.9	0.1	23,100	11.4	0.01	87.3	0.03	87.3	0.4

Low secondary effluent alkalinity conditions in September and October 1970 limited the alum dosage that could be fed to the tertiary system and still maintain a pH condition compatible with alum floc formation. The chemical metering pump assigned to add polyelectrolyte between the tertiary tube settler and mixed-media filter was used to feed soda ash (sodium carbonate) to the tertiary influent, as a supplemental source of alkalinity.

Polyelectrolyte could not be added prior to filtration, until a fourth metering pump was installed on 7 April 1971.

Polymer dosages in the tertiary influent (presettler) were varied from 0.4 to 3.5 mg/l. Polymer dosages introduced immediately ahead of filtration (prefilter) ranged from 0.02 to 0.22 mg/l. Figure 31 shows the feed rates of the various chemicals added to the tertiary system during the demonstration program.

FILTRATION RATE

The tertiary system was operated within four flow rate ranges during various periods of the demonstration program, resulting in the following filter surface hydraulic loadings:

1.4 - 1.8 gpm/ft^2 (70 - 90 gpm)

2.2 - 2.3 gpm/ft^2 (110 - 115 gpm)

2.8 - 3.2 gpm/ft^2 (140 - 160 gpm)

3.6 - 4.0 gpm/ft^2 (180 - 200 gpm)

The tertiary flow rates and corresponding time periods during the project are shown on Figure 32.

In general the filtration rate did not appear to substantially influence the throughput volume per filter cycle or effluent quality. The wide variation in the secondary effluent characteristics and the range of alum and polyelectrolyte dosages applied allowed only a qualitative assessment to be made, however.

The total volume filtered per run at rates of 2.2 - 2.3 gpm/ft^2 and 3.6 - 4.0 gpm/ft^2 are listed in Table 11 for similar suspended solids, pH and chemical feed conditions.

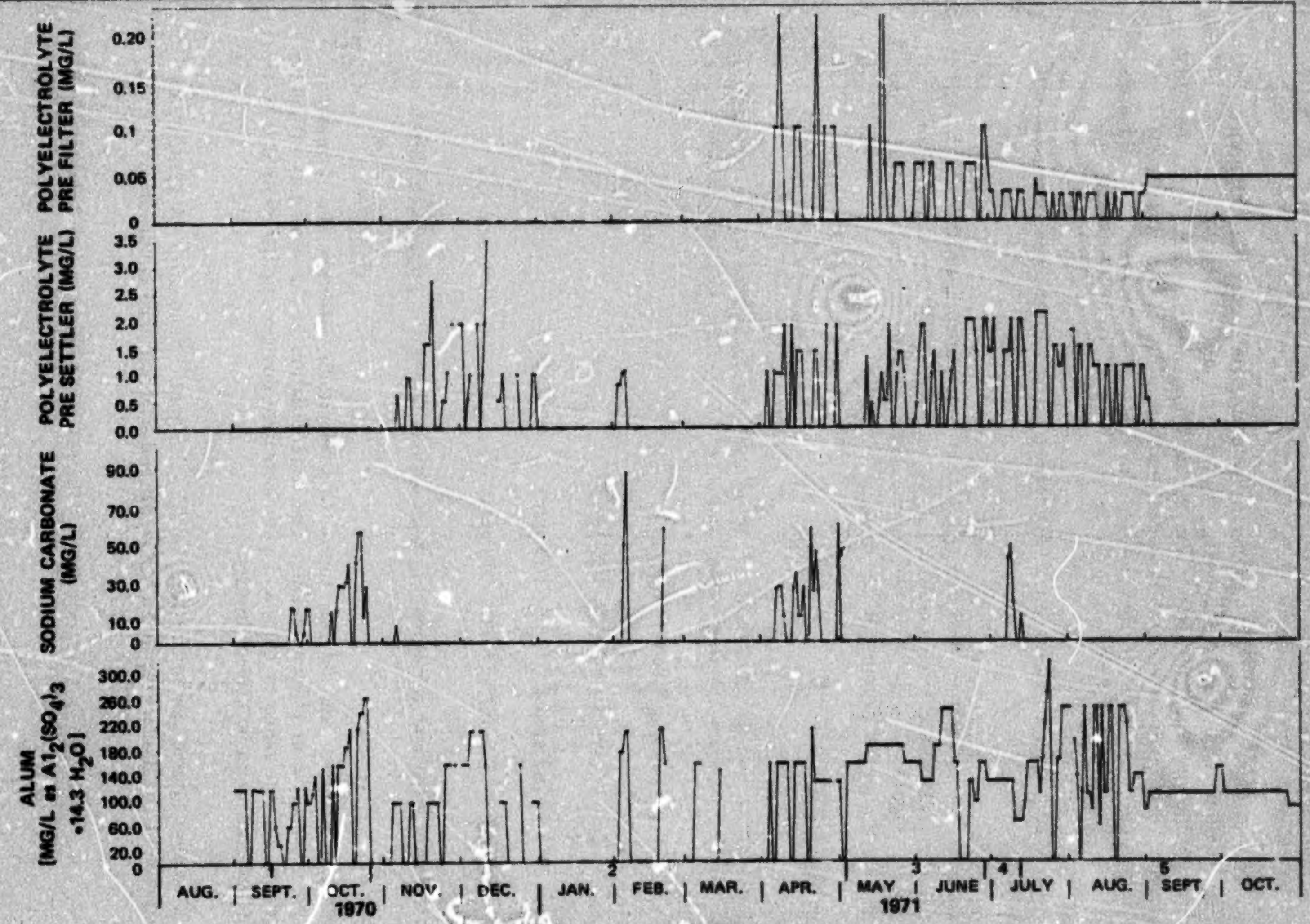


FIGURE 31
TERTIARY CHEMICAL DOSAGES VERSUS TIME

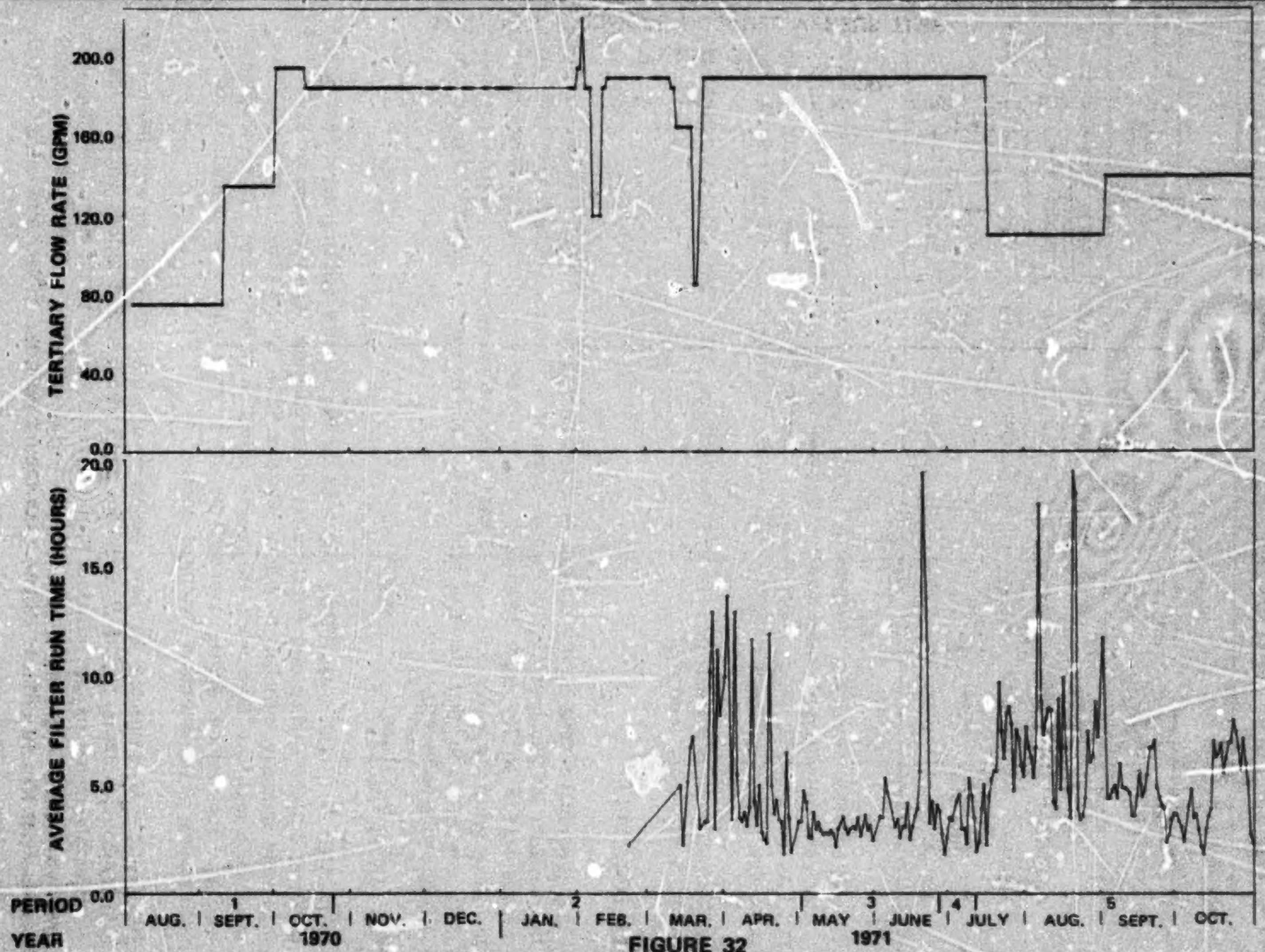


FIGURE 32
TERTIARY FLOW RATE AND FILTER RUN TIME VERSUS TIME

TABLE 11

FILTRATION RATE COMPARISON

FILTRATION RATE (GPM/FT^2)	2.2–2.3				3.6–4.0			
ALUM FEED (MG/L)	AVG. VOL. FILTERED (GAL/RUN)	SEC. EFF. TSS (MG/L)	TER. EFF.		AVG. VOL. FILTERED (GAL/RUN)	SEC. EFF. TSS (MG/L)	TER. EFF	
			TSS (MG/L)	pH			TSS (MG/L)	pH
0	118,800	92	11	7.2	124,300	43–92	6–15	6.8–7.5
126–135	44,600	61	2	6.4	46,000	31–63	1–4	6.1–6.5
155–160	44,700	34–60	1–5	6.5–6.7	39,800	45–51	1–3	6.4–6.8
182–190	35,400	67	1	6.25	33,600	29–66	1–3	6.0–6.3
240–242	34,300	70–78	1	6.5–6.6	35,100	39–75	1	6.1

The tertiary effluent quality depended primarily on the characteristics of the secondary effluent and the chemical dosages applied, and did not appear to be affected by filtration rate. As an example, at an alum dosage of 155 to 160 mg/l, the tertiary effluent suspended solids averaged 1.2 mg/l at a flow rate of 110 gpm and 1.6 mg/l at 190 gpm.

Additional study is needed to verify these observations, where the influent characteristics can be more closely controlled.

FILTER CYCLE

Filter run times ranged from a maximum of 106 hours at a flow rate of 75 gpm without the addition of chemicals to less than 2 hours at a flow rate of 190 gpm and alum dosages of 155 to 242 mg/l. The length of filter cycle was affected by several factors, including solids carryover from the tertiary tube settler, secondary effluent suspended solids concentration, and chemical feed rates. The influence of each of these factors is discussed in Section VIII.

EFFLUENT QUALITY

BOD AND TSS–The total BOD_5 and TSS in the tertiary effluent depended on the chemical dosages applied and effectiveness of coagulation. Alum dosages of 80 to 105 mg/l were generally sufficient to reduce the total BOD_5 to less than 10 mg/l and TSS to less than 5 mg/l. At these levels, the overall reduction of BOD_5 and TSS approached 99 percent. The soluble BOD_5 averaged about 6 mg/l and was dependent on the performance of the secondary treatment system.

Nitrification is suspected to have occurred in some of the tertiary BOD tests. Simultaneous oxidation of the nitrogeneous and carbonaceous fractions would produce a considerably higher oxygen demand than the carbonaceous phase alone, resulting in an apparent high BOD_5. During periods when the secondary effluent was completely nitrified and tertiary effluent TSS was less than 3 mg/l, the tertiary effluent total BOD_5 averaged about 5 mg/l.

Tertiary total and soluble BOD_5 and TSS are shown versus time on Figure 33.

NITROGEN AND PHOSPHORUS—As would be expected, the concentration of ammonium, nitrite, and nitrate in the secondary and tertiary effluent were essentially identical. The residual organic nitrogen (total kjeldahl less ammonia nitrogen) depended on the residual TSS of the secondary effluent.

The concentration of orthophosphate remaining in the plant effluent after tertiary treatment was a function of the secondary orthophosphate concentration, alum dosage, and tertiary pH. The relationship of these three variables on orthophosphate removal is discussed in Section VIII.

Orthophosphate residuals of less than 0.5 mg/l (as P) could be maintained with alum dosages of 100 to 200 mg/l. Under optimum pH conditions, orthophosphate concentrations of less than 0.1 mg/l (as P) were achieved.

Total phosphate residuals depended on the efficiency of TSS removal in the tertiary system, as well as the secondary orthophosphate concentrations, alum dosage, and pH. At tertiary TSS levels of 1-3 mg/l, the ortho and total phosphate concentrations were essentially the same. The average total phosphate removal through the plant at alum dosages above 125 mg/l exceeded 95 percent.

The concentration of total nitrogen, ammonium, nitrate, orthophosphate and total phosphate in the tertiary effluent versus time are shown on Figure 34.

pH, ALKALINITY AND TURBIDITY—pH (and its relationship with total alkalinity) was perhaps the single most important variable to the control and performance of the tertiary system. The alkalinity of the secondary effluent and pH of the tertiary effluent affected alum coagulation, phosphate and TSS removal, and turbidity level of the tertiary effluent. Tertiary effluent pH, total alkalinity and turbidity are plotted versus time on Figure 35.

The alkalinity equivalent of the alum $[Al_2(SO_4)_3 \cdot 14.3\ H_2O]$ used during the project was determined experimentally to be about 0.36 mg/l equivalent $CaCO_3$ per mg/l of alum. In order for a chemical floc to develop, the secondary effluent alkalinity had to be sufficient to offset the acidity of the alum dosage and maintain the pH above 5.0. If the secondary effluent alkalinity was not adequate, sodium carbonate was added to the tertiary influent to control pH.

The base turbidity on Figure 35 was the average turbidity of the tertiary effluent. The peak turbidity was the level recorded at the end of the filter run and indicates the magnitude of turbidity breakthrough.

The lowest residual TSS and turbidity levels occurred in a pH range of 5.8 to 6.3, as shown on Figure 36. The data plotted on Figure 36 represent the average of the observed values at a given pH for alum dosages of 105 to 240 mg/l. This pH range is believed to be the optimum zone for alum coagulation of the secondary effluent. Under these pH conditions, negatively charged colloidal material suspended in the wastewater is

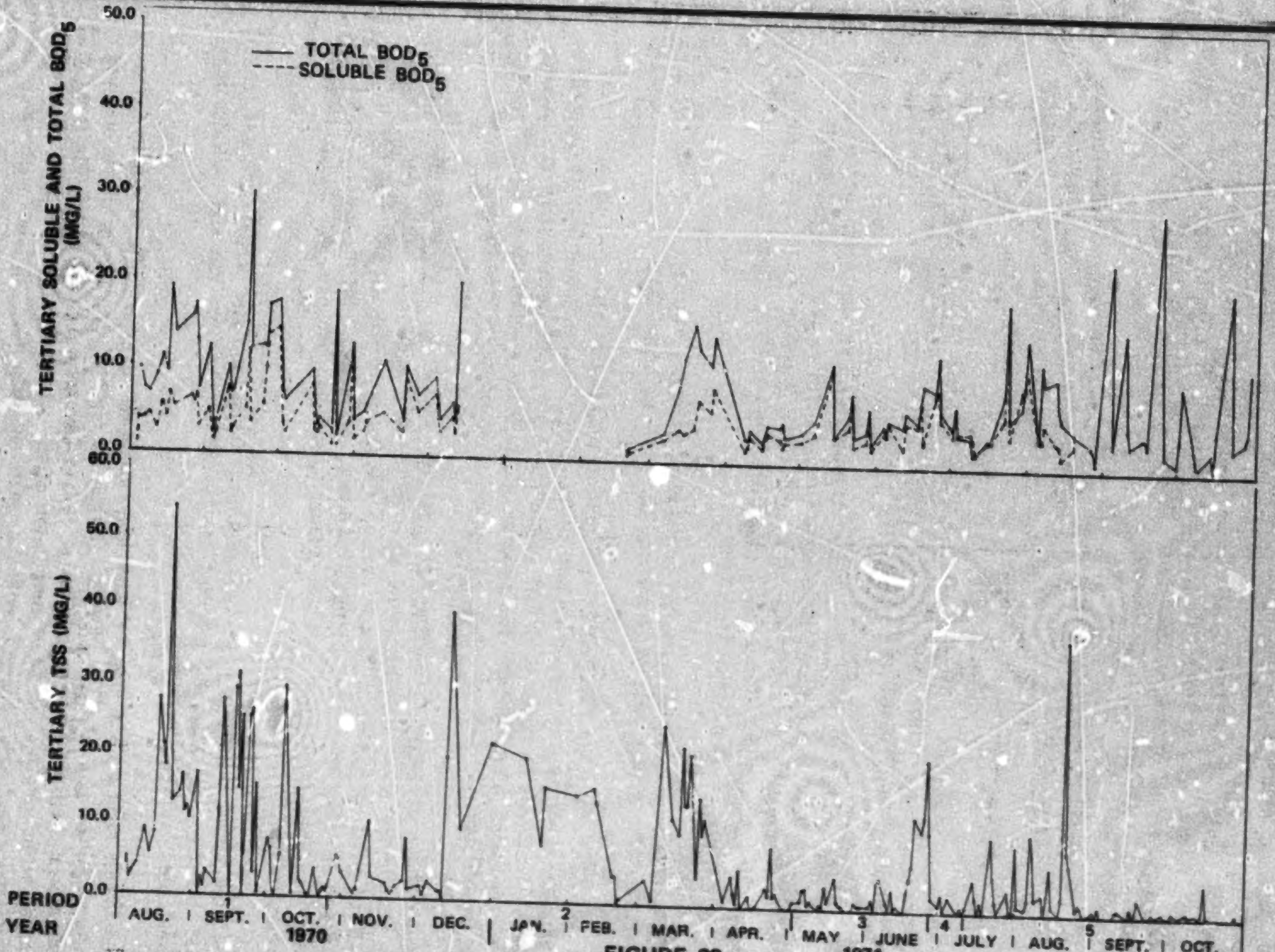


FIGURE 33
TERTIARY EFFLUENT TOTAL AND SOLUBLE BOD AND TSS CONCENTRATION VERSUS TIME

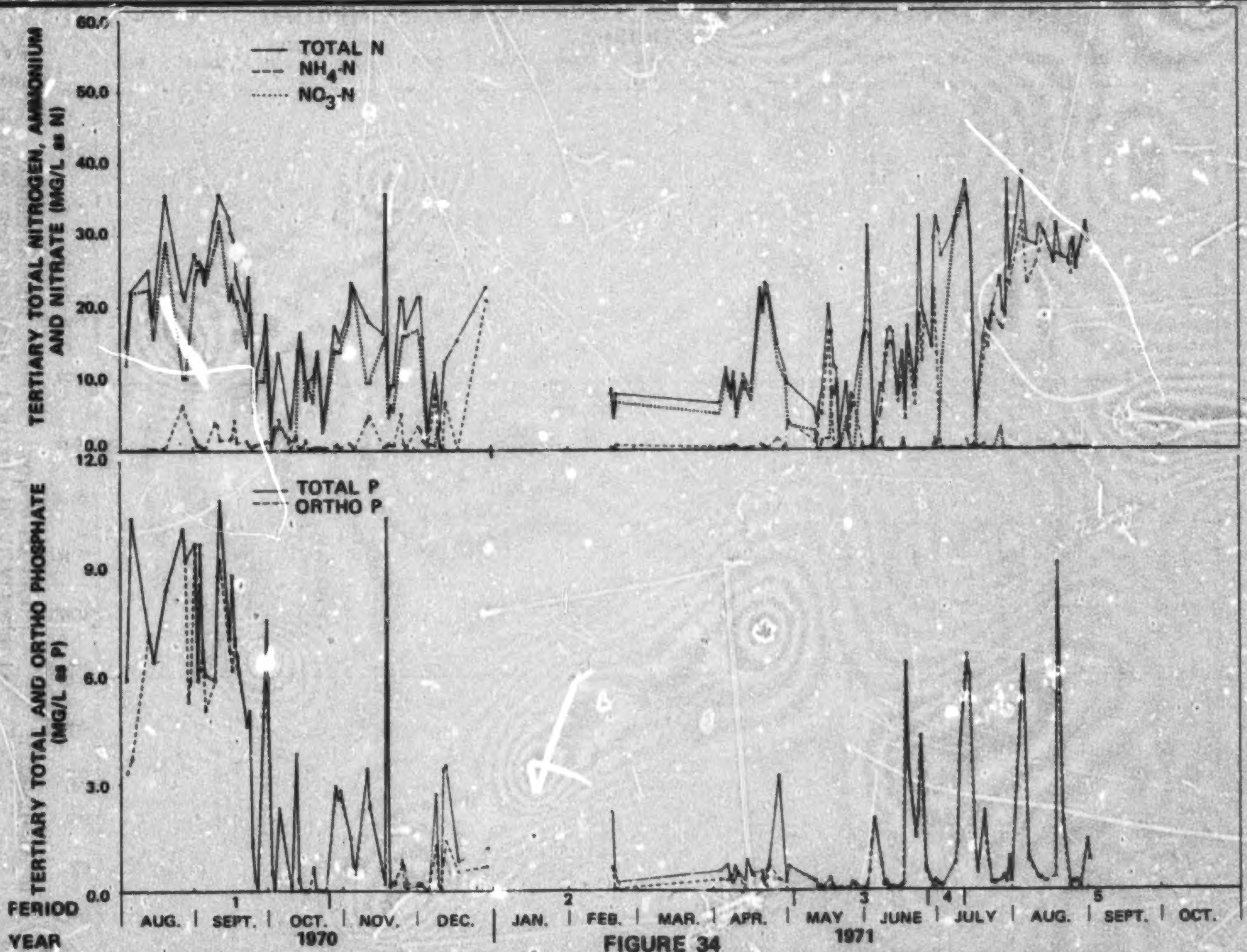


FIGURE 34

TERTIARY EFFLUENT NITROGEN AND PHOSPHOROUS CONCENTRATIONS VERSUS TIME

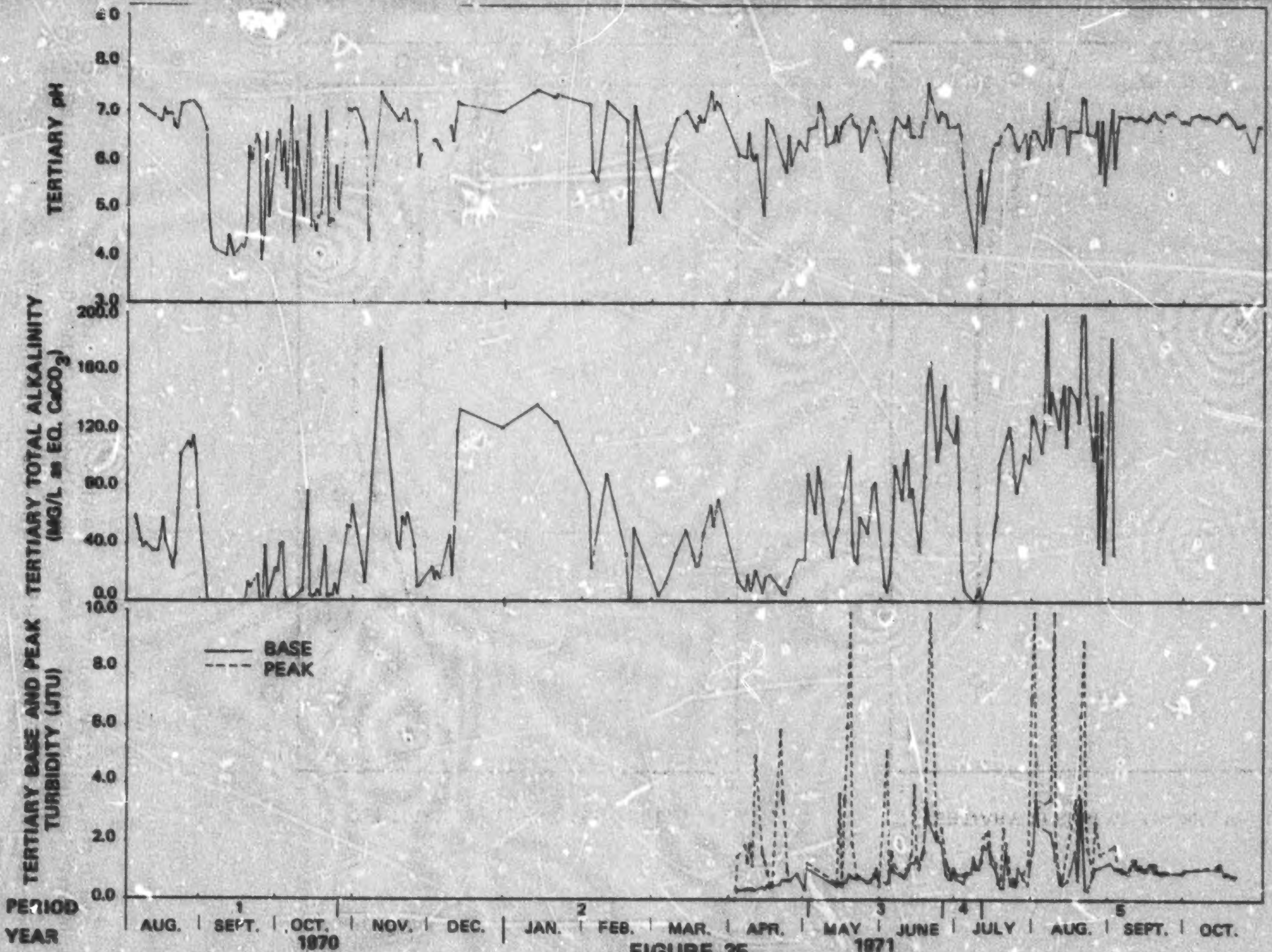


FIGURE 35
TERTIARY EFFLUENT pH, ALKALINITY AND TURBIDITY VERSUS TIME

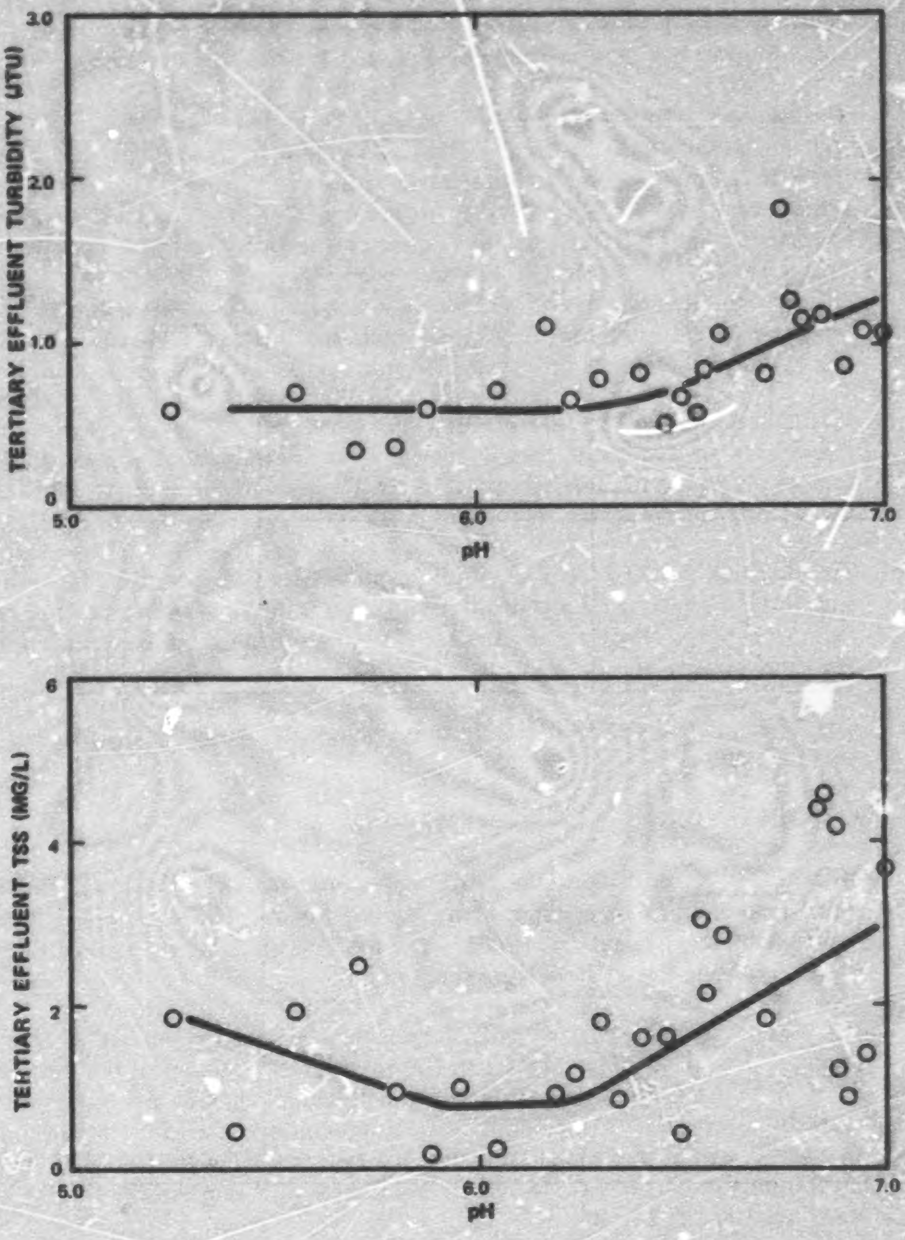


FIGURE 36

TERTIARY EFFLUENT TSS AND TURBIDITY VERSUS pH neutralized by positively charged hydrated aluminum ions. The result is a destabilization and entrapment of the suspended solids, as the alum floc develops.

The tertiary effluent turbidity also correlated with the nitrate level in the tertiary effluent as shown on Figure 37. This correlation is, at least partially, a reflection of the pH and alkalinity conditions resulting from nitrification. The apparent relationship of nitrate concentration to turbidity is presented here to illustrate the affect of secondary treatment conditions, such as nitrification, on tertiary treatment system performance.

Nitrate ion may also have a possible secondary effect on chemical coagulation. As a potential determining ion, it is conceivable that nitrate could reduce the zeta potential of the hydrated aluminum colloid and promote coagulation.

POLYELECTROLYTE ADDITION–The affect of adding polyelectrolyte ahead of the flocculator to assist floc formation was difficult to assess. In general, it did not appear that applying polyelectrolyte at this point enhanced chemical floc formation, improved the settling characteristics of the floc, or lengthened the filter run time.

The addition of 0.02 to 0.04 mg/l of polyelectrolyte between the tertiary tube settler and filter effectively controlled turbidity breakthrough. At dosages above 0.06 to 0.10 mg/l, the polyelectrolyte appeared to "bind up" the filter bed and appreciably shorten the filter cycle.

The effect of polymer on the tertiary treatment process is discussed in more detail in Section VIII.

CHLORINATION

MPN–Plant effluent chlorination data are summarized in Table 12. Because of the very limited amount of data obtained during the demonstration program, Table 12 has been supplemented with data obtained from the Oregon State Department of Environmental Quality collected after termination of the project.

A chlorine residual of 1.2 mg/l or greater was adequate to reduce the total coliform index to less than 1,000 MPM per 100 ml after a minimum of 0.9 hours contact time in 5 of 6 samples analyzed. Water of this bacterial quality is considered acceptable for recreational purposes, including such water-contact activities as swimming and water skiing, by several governmental regulatory agencies including the Ohio River Valley Water Sanitation Commission (ORSANCO) [11]. In 4 of 5 samples the fecal coliform MPN was less than 100 per 100 ml.

The sample collected on 15 May 1972 had an unusually high coliform count, even though the chlorine residual was 2.5 mg/l. The chlorine residual measurement is suspected to have been in error, considering the small dosage of chlorine applied. The high chlorine residual measurement, however, may have resulted from much of the chlorine demand being exerted by ammonium ion to form chloramines with little or no free chlorine residual available for disinfection. The difference in plant influent and secondary effluent alkalinity indicates that the secondary treatment system may not have been completely nitrified. A considerable amount of ammonia likely remained in the secondary effluent.

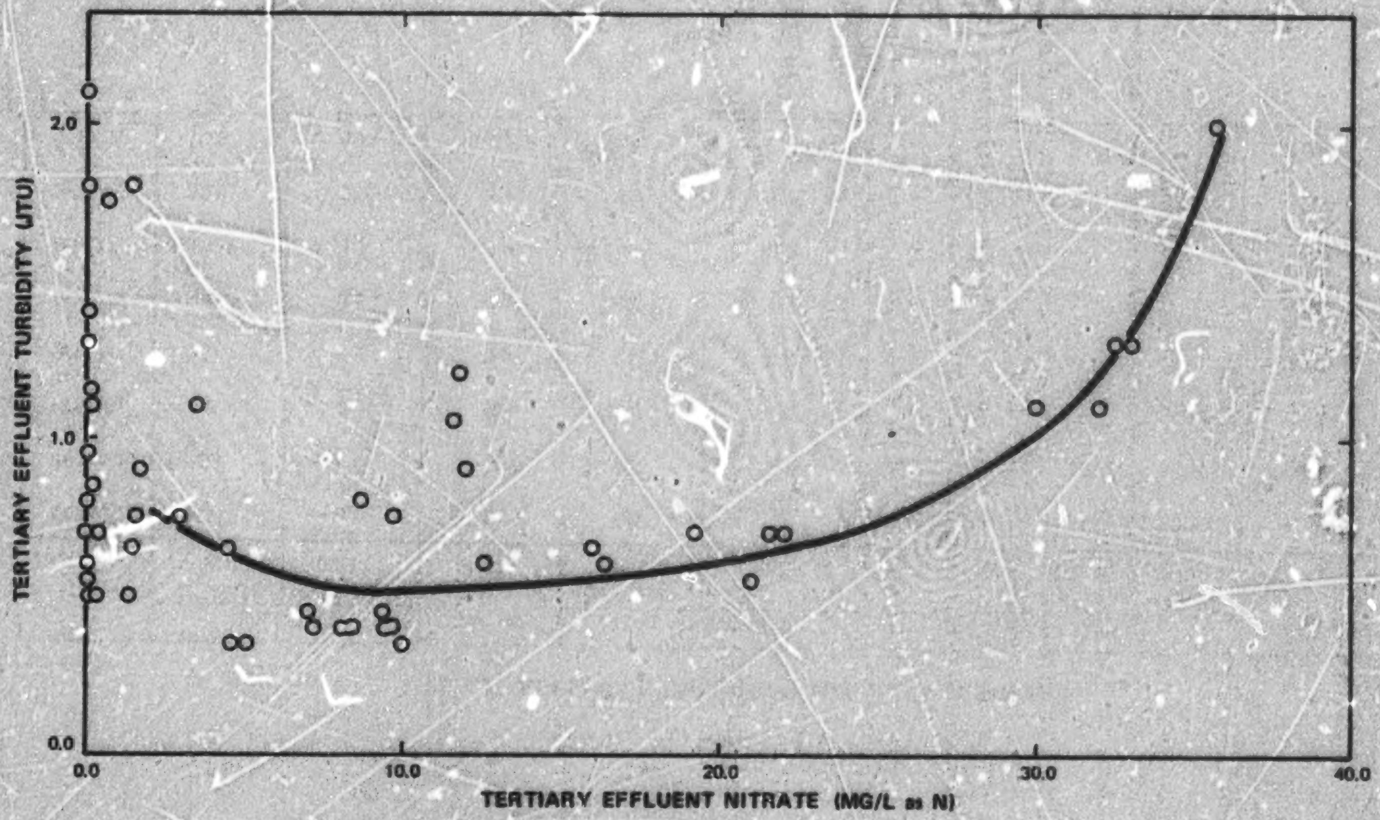


FIGURE 37

TERTIARY EFFLUENT TURBIDITY VERSUS NITRATE CONCENTRATION

TABLE 12

PLANT EFFLUENT CHLORINATION DATA SUMMARY

DATE	CHLORINE		CONTACT TIME (HRS.)	TOTAL ALKALINITY		TERTIARY EFFLUENT				COLIFORM (MPN/100 ML)	
	DOSAGE (MG/L)	RESIDUAL (MG/L)		INFLUENT (MG/L)	SEC. EFF. (MG/L)	BOD	TSS	KJEL-N (MG/L)	NH_3-N (MG/L)	TOTAL	FECAL
8/12/70	12.	0.6	2.2	180	35	11.2	5.8	2.4	0.39	<45	<45
10/27/70	5.	0	0.9	166	53	10.3	5.6	2.4	0.62	24,000	2,300
7/15/71	14.	2.3	0.9	206	136	–	1.6	1.9	0.665	8	–
9/13/71	7.	1.8	1.2	196	199	–	0	–	–	<45	<45
5/15/72	5.5	2.5	1.2	200	121	3.0	8	–	–	1,300	1,300
6/6/72	9.5	1.2	1.2	180	57	1.4	1.0	–	–	60	<45
6/20/72	39.5	2.3	1.2	167	38	–	2.0	–	–	230	60
7/17/72	5.5	0.15	1.2	200	110	–	3.0	–	–	7,000	2,100

Conversion of ammonia to nitrate in the secondary system will substantially reduce the chlorine dosage required to reach the breakpoint and produce a free chlorine residual.

With complete nitrification occurring in the secondary treatment system, a chlorine dosage of about 10 mg/l or less was adequate to maintain a chlorine residual above 1.0 mg/l.

BREAKPOINT CHLORINATION TESTS—Three breakpoint chlorination tests were run during June and July 1971. The results of the tests indicated breakpoint demands ranging from 125 to 206 mg/l. During this period, the ammonium concentration varied between 14 to 23 mg/l (as N), which undoubtedly increased the demand above that expected when the secondary treatment system was completely nitrified. Difficulties in preparing and standardizing reagents, however, make the data subject to question.

WASTE SOLIDS STORAGE

WASTE ACTIVATED SLUDGE STORAGE

GENERAL—Waste activated sludge (WAS) from the aeration-surge basin was stored in two adjacent 42,000 gallon holding lagoons (Figure 10). WAS Lagoon No. 2 was examined at the conclusion of the demonstration program to:

1. Determine the suitability of lagoons for storage of WAS from an extended aeration system.
2. Estimate the volume and assess the feasibility of dewatering WAS in the lagoons.
3. Characterize the solids after dewatering and estimate the loss of volatile solids as a result of anaerobic decomposition.
4. Estimate cleaning frequency and recommend a method of disposal.

A record of the solids wasted to WAS Lagoon No. 2 was maintained during the demonstration program. On 9 June 1971, the supernatant was decanted to the sludge blanket. The sludge deposit was allowed to dry through the summer and the depth of the sludge deposit measured at varying time intervals. A profile was made of the sludge deposit on 16 September 1971 to determine the quantity and characteristics of solids remaining in the lagoon.

No objectionable odors were detected from the storage lagoons at any time during the project. During the dewatering operation, when a crust was beginning to form, a slight "earthy," but unobjectionable, odor was observed when standing 10-15 feet downwind of the lagoon.

DEWATERING—Figure 38 shows the depth of the sludge deposit with time as the material dewatered and dried. The depth of the sludge in the storage lagoon, measured from the base of the outlet structure, decreased from 25 inches to 11-1/2 inches over a period of approximately nine weeks. The ground water level is suspected to have prevented the sludge from continuing to dewater.

As the sludge dewatered, a brown crust developed over the deposit. Within 3-4 weeks the crust cracked and dried to a cinder-like material. Below the crust, the black sludge had the consistency and appearance of thick chocolate pudding.

Photographs of the sludge storage lagoon taken at the time the lagoon was profiled are contained in Appendix B.

SLUDGE CHARACTERISTICS—While the sludge deposit was being profiled, samples were taken from five locations in the lagoon at various depths to determine the total solids and total volatile solids content of the material. The results of these analyses are summarized on Figure 39.

The material had an average total solids content of 18 percent. The solids were found to be progressively more moist with depth, except for the bottom three inches where the solids concentration appeared to increase. The volatile fraction of the solids averaged 52 percent. In the upper 10 inches the total volatile solids ranged between 54 to 60 percent. Accumulation of inorganic materials in the bottom three inches reduced the volatile fraction to around 34 percent.

SOLIDS ACCUMULATION—From plant records, approximately 25,000 pounds of total (dry) solids were wasted to WAS Storage Lagoon No. 2. The profile of the sludge deposit indicated an accumulation of approximately 57.2 cu yds at an average depth of about 13 inches. The bulk density was calculated to be 66 pounds per cubic foot assuming a specific gravity of 1.4 for the dry sludge solids [11]. At this bulk density the quantity of solids remaining was calculated to be 18,300 pounds. Approximately 27 percent of the WAS was either destroyed by anaerobic decomposition, or lost as suspended solids in the return sludge lagoon supernatant.

For comparative purposes, an attempt was made to balance the WAS with solids accumulation in the lagoon based on ash content. The volatile content of the WAS for a limited number of analyses performed in August 1971, averaged 73.5 percent. Assuming this value was representative of the WAS over the entire demonstration program, the inorganic fraction of the total solids wasted was estimated to be 6,625 pounds. At an average ash content of 43.5 percent, the solids accumulation in WAS Storage Lagoon No. 2 contained approximately 8,000 pounds of inorganic material.

Comparing these estimates, it appears that either more solids were wasted to the storage lagoons than recorded or the average volatile fraction of the WAS was closer to 68 percent.

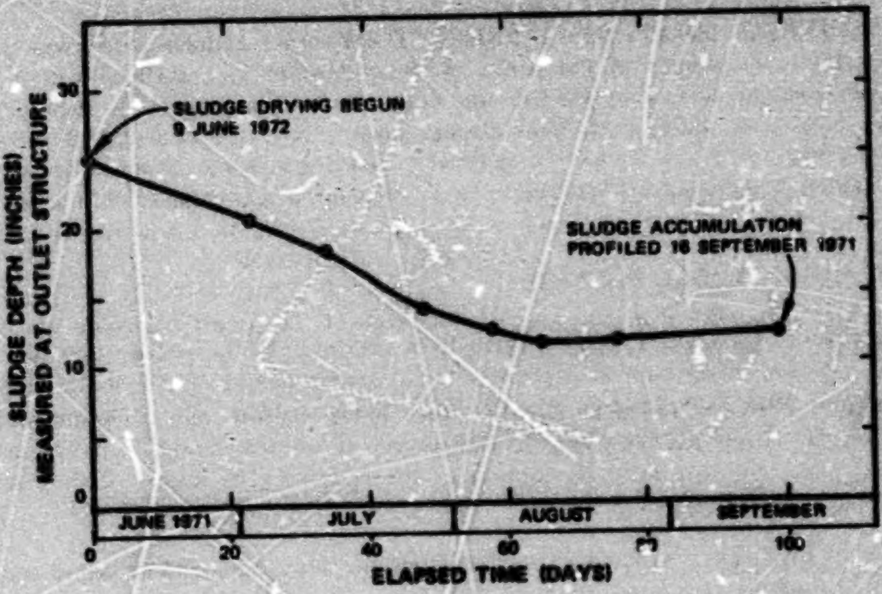


FIGURE 38

WASTE ACTIVATED SLUDGE STORAGE LAGOON DRYING VERSUS TIME

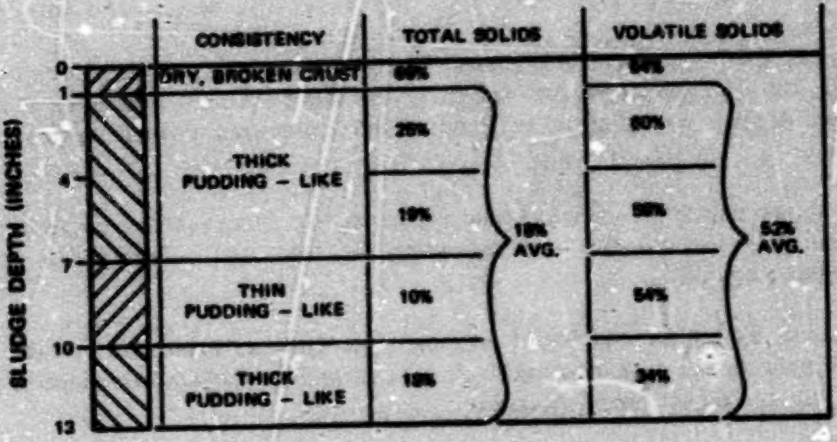

SLUDGE DEPTH (INCHES)
0
1
4
7
10
13
CONSISTENCY
TOTAL SOLIDS
VOLATILE SOLIDS
DRY, BROKEN CRUST
66%
64%
THICK PUDDING – LIKE
25%
60%
19%
58%
THIN PUDDING – LIKE
10%
54%
THICK PUDDING – LIKE
18%
34%
18% AVG.
52% AVG.

FIGURE 39

WASTE ACTIVATED SLUDGE STORAGE LAGOON PROFILE

CLEANING FREQUENCY–At the plant design organic loading, removal of the solids from both WAS sludge storage lagoons will be required at 9 to 12 month intervals.

ADDITIONAL SOLIDS DEWATERING STUDY–An additional study was conducted to determine the ability of the stored WAS to dewater in a shallow bed. A drying bed approximately 20 feet by 80 feet was improvised to the west of the aeration-surge basin and filled with sludge from WAS Storage Lagoon No. 1 to an average depth of 6 inches. The sludge dried to a cinder-like material in about 4-6 weeks. Analyses of the dried solids indicated a moisture content of 9 to 12 percent and a total volatile solids content of 50-54 percent.

CHEMICAL SLUDGE STORAGE

GENERAL–Chemical sludge resulting from backwash of the tertiary unit was pumped to the chemical sludge storage lagoons after a settling period of 2-4 hours in the chemical sludge holding tank (Figure 10). Chemical Sludge Lagoon No. 2 was examined at the conclusion of the demonstration program to:

1. Determine the suitability of lagoons for storing chemical sludge.
2. Estimate the volume and determine the dewatering and drying characteristics of the accumulated solids.
3. Characterize the sludge after storage and dewatering.
4. Estimate cleaning frequency and recommend a method of disposal.

Except for short periods in October 1970, January, March, and May 1971, totaling approximately 9 weeks, all of the chemical sludge produced from plant startup through 21 June 1971 was stored in Chemical Sludge Lagoon No. 2. During the last week of June 1971, the supernatant was decanted to the sludge blanket. Beginning 2 July 1971 the sludge deposit was allowed to dewater by evaporation through the remainder of the summer.

The sludge deposit in Chemical Sludge Lagoon No. 2 was profiled on 16 September 1971 to calculate the quantity and determine the total and volatile solids content of the accumulated solids.

Chemical Sludge Lagoon No. 1 was used for temporary influent storage while repairs were made to the aeration-surge basin and secondary clarifier. This lagoon was also used for WAS storage on several occasions.

During the fall of 1970 and again in the spring of 1971, heavy algae blooms developed in both chemical storage lagoons, particularly Chemical Sludge Lagoon No. 2. Algae in the supernatant return from the lagoons passed through the aeration-surge basin and could be seen in the secondary and tertiary effluent suspended solids determinations during April and May 1971. The fraction of the suspended solids attributable to algae was not determined, however. The alkalinity and pH of the return supernatant followed a predictable pattern as a result of algal activity. At the peak of the algae bloom in April 1971, the alkalinity was 58 and pH 8.8. As the bloom dissipated during the third week in May, the pH dropped to 6.85 and alkalinity increased to 91.

Photographs of Chemical Sludge Storage Lagoon No. 2, taken at the time the lagoon was profiled, are contained in Appendix B.

SOLIDS DEWATERING–The change in depth of the sludge deposit in Chemical Sludge Lagoon No. 2 with time, as the lagoon dewatered, is shown on Figure 40. The depth, measured from the base of the outlet structure, decreased from 26 inches to 10-1/2 inches over a period of approximately 8 weeks. Ground water prevented the material from concentrating further and drying.

A floating mat 1/2- to 1-inch thick developed over about 80 percent of the surface of the lagoon, as the sludge liquor concentrated. No objectionable odors were generated from the lagoon, even during the dewatering period.

SOLIDS CHARACTERISTICS–Composite samples were taken from eight locations in Chemical Sludge Lagoon No. 2. The total (dry) solids content of these samples average 4.0 percent, ranging between 2.9 to 5.1 percent. The volatile fraction of the dry solids averaged 48.5 percent and varied from 40.8 to 51.8 percent. The bulk density of the material averaged 59.8 pounds per cubic foot. The sludge was slightly more concentrated around the influent pipe in the center of the lagoon and in front of the outlet structure.

The floating mat had a total solids content of 13-14.5 percent. The volatile content of this material was about the same as the composite chemical sludge samples.

SOLIDS ACCUMULATION–The total amount of chemical sludge discharged to Chemical Sludge Lagoon No. 2 was calculated to be 30,300 pounds, including both alum and organic solids. From the profile made of the sludge accumulation, the quantity of solids remaining in the lagoon was calculated to be 21,200 pounds. The 30 percent reduction is attributed to anaerobic decomposition and loss of suspended solids in the supernatant return.

An attempt was made to balance the inert fractions of the chemical sludge wasted to the lagoon, the suspended solids lost in the lagoon supernatant return, and the solids accumulation in the lagoon. The volatile content of the organic solids removed by the tertiary system was assumed to average 73.5 percent, based on mixed liquor volatile solids analyses. Sludge samples collected when the pond was profiled, had an average volatile content of 48.5 percent. The suspended solids in the lagoon supernatant were assumed to be 75 percent volatile.

Based on these data, the inert fraction of the chemical sludge discharged to the lagoon less the inert fraction of the suspended solids in the supernatant return was calculated to be 12,300 pounds. The inert fraction of the solids retained by the lagoon was calculated to be 10,900 pounds.

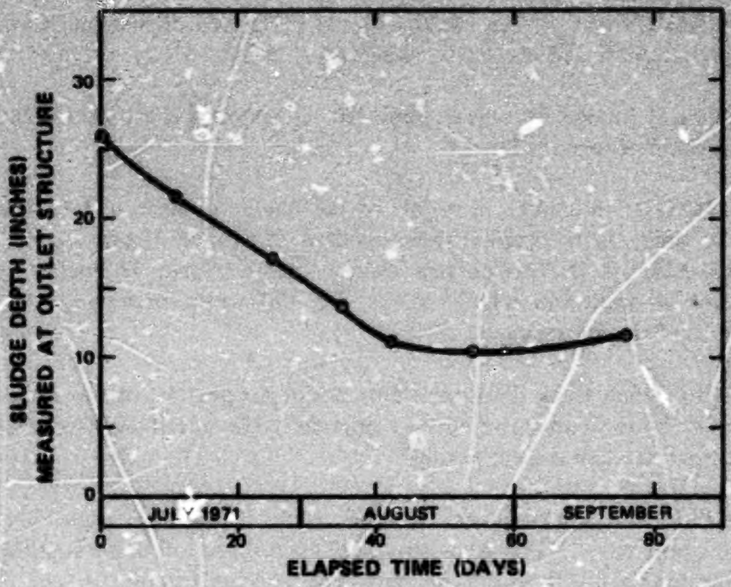


FIGURE 40

CHEMICAL SLUDGE LAGOON DRYING VERSUS TIME

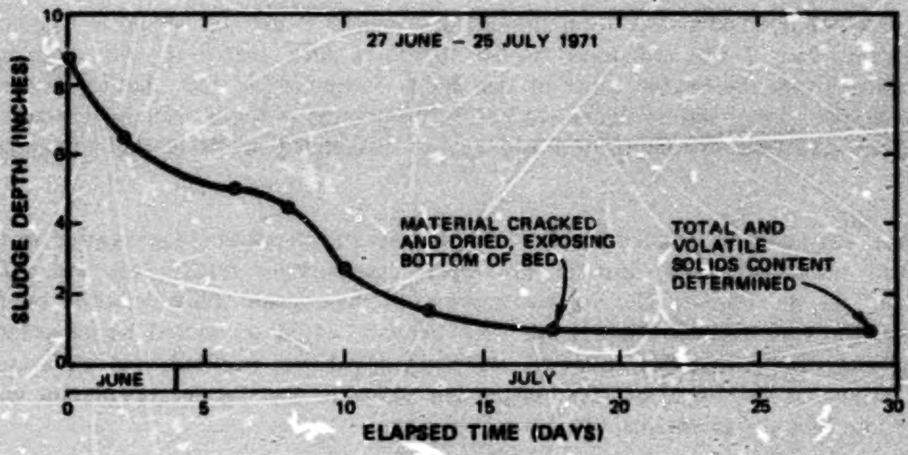


FIGURE 41

CHEMICAL SLUDGE ALIQUOT DRYING VERSUS TIME

The agreement of these estimates to within 13% tends to add validity to the calculated quantities of total chemical sludge wasted to and accumulated by the chemical sludge storage lagoon during the project.

CLEANING FREQUENCY—At a design flow of 0.28 mgd, removal of solids from each lagoon will be required at 2 to 3 year intervals.

ADDITIONAL CHEMICAL SLUDGE DEWATERING STUDY—A small drying bed (3 ft x 3 ft x 1 ft) was constructed on the dike of Chemical Sludge Lagoon No. 2 to determine the drying characteristics for a small aliquot of the chemical sludge, without the influence of ground water. On 27 July 1971, the bed was filled to a depth of 8-3/4 inches with lagoon sludge having a dry solids content of 3.3 percent. Within 17 days the depth of the bed was reduced to about 1 inch, at which point the material cracked and dried to a whitish-gray cinder-like solid. After 27 days, the total solids content of the dried solids was determined to be approximately 68 percent. The volatile fraction of the dry solids averaged 45 percent.

The results of the aliquot drying study are shown on Figure 41. Photographs of the drying bed at the end of the test period are contained in Appendix B.

COMBINED RETURN SLUDGE LAGOON SUPERNATANT

TSS, BOD and pH analyses were performed on samples of the combined return sludge lagoon supernatant collected from the discharge to the plant influent pump station. The results of these analyses are listed in Table 13 below:

TABLE 13
COMBINED RETURN SLUDGE LAGOON
SUPERNATANT CHARACTERISTICS

	Average	Maximum	Minimum
TSS (mg/l)	73	160	30
BOD (mg/l)	51	96	26
pH (mg/l)	6.99	7.55	6.5

OPERATIONAL CONSIDERATIONS

A number of problems were encountered during the demonstration program which affected plant performance and the overall conduct of the project. The more significant problems are described in the following paragraphs.

OPERATOR ACCIDENT

Two weeks after plant startup in April 1970, the treatment plant operator/research technician was seriously injured. He was unable to return to work on a full-time basis until the latter part of July 1970. A temporary operator was employed part-time to operate and maintain the plant. His analytical experience was very limited, however, and the scheduled sampling and testing program was suspended until the original operator was able to resume work.

PVC LINER FAILURE

On 8 May 1970 the PVC liner in the east end of the aeration-surge basin floated as the result of gas buildup under the liner. Inspection revealed that the liner had pulled away from the footing, allowing seepage under the membrane. Subsequent septic conditions and gassification caused the liner to float. The liner was repaired and reattached to the aeration-surge basin ringwall footing.

On 22 June 1970 the liner was again found to be floating on the surface in the same location of the aeration-surge basin. The liner was carefully inspected around the entire perimeter of the basin. In addition to finding the liner pulled away from the ringwall footing at several locations along the east end of the basin, a small cross-hatched cut was detected in the liner. The cut had apparently been made to relieve entrapped air when the liner was originally installed, but was not repaired.

It is suspected that mixed liquor had slowly leaked through this cut and under the liner. Decomposition of organic material provided sufficient gas pressure to tear the liner from the footing. This allowed more liquid to seep under the liner and accelerated gassification.

When the liner was repaired for the second time, a gas relief system was also installed to allow any gas forming under the liner to escape without damage to the membrane.

The second liner repair was completed in mid-July 1970. No further problems with the PVC liner were encountered during the remainder of the demonstration programs. Pictures of the liner repair and gas relief system are included in Appendix B.

INFLUENT PUMPS

The submersible influent pumps jammed with rags and other debris on an average of 2-3 times each month. The pump had to be raised with the aid of a backhoe and the impellers cleaned each time. Though not a difficult task, the periodic pump cleaning was a considerable nuisance.

AERATION EQUIPMENT

Shortly after startup of the plant, the aerators began to periodically trip the thermal overload circuit breakers, due to excessive current draw. The aerators were examined but no evidence of debris binding or clogging the propellers could be found.

Amperage recorders were placed on the power feed lines. A near linear increase in current draw was observed over time periods lasting several hours to several days from the time units were activated, until excessive current draw actuated the circuit breakers.

In March 1971, the aerators were pulled from the basin. The manufacturer changed the pitch of the propellers in an attempt to reduce the current draw. The modification did not completely resolve the problem. The aerators continued to kick off periodically, especially during warm weather.

SODA ASH FEED

Periods of complete nitrification substantially reduced the aeration-surge basin alkalinity. Under these conditions, soda ash was used, as necessary, to supplement the secondary effluent alkalinity to maintain pH control in the tertiary system. A soda ash feed system was devised using one of the polyelectrolyte metering pumps and the existing chemical mixing and storage tankage.

SECONDARY TUBE CLARIFIER MODIFICATIONS

Solids accumulation in the secondary clarifier tubes had a pronounced effect on the secondary effluent quality in the first half of the demonstration program. On 4 October 1970 gassification of the entrapped solids provided sufficient bouyancy to float the tube modules in the clarifier structure. Restraining bands were added to hold the tube modules in place.

In December 1970, MicroFLOC Corporation tested an air sparging system designed to clean and prevent solids from collecting in the tube bundles of the secondary clarifier. The tests were successful and a sparging system was installed during February and March 1971. Daily sparging of the tubes for a period of 15 minutes considerably improved the performance of the unit.

At the same time the sparging system was being installed, the holes in the clarifier draw-off pipe were enlarged. The reduction of the head loss in the tertiary influent line, as a result of this modification, tended to correct an air entrainment problem,which caused flow surges through tertiary influent pump.

FLOCCULATOR MODIFICATIONS

Air entrainment in the discharge from the overflow box to the flocculator tank caused severe hydraulic short circuiting. An adjustable constrictive collar was placed in the overflow box discharge line to maintain a constant water level over the discharge pipe and prevent air from being drawn into the flocculator.

Sodium chloride tracer studies conducted by MicroFLOC after the constrictive collar was put in, however, indicated the actual mean detention time of the flocculator tank to be about 50 percent of the theoretical value. To further reduce hydraulic short circuiting, a horizontal baffle was installed in the flocculator tank in January 1971. Subsequent tracer studies indicated this modification increased the actual mean detention time to approximately 75-80 percent of the theoretical value.

TERTIARY TUBE SETTLER/FILTER MODIFICATIONS

It appeared from visual observations that the chemical floc entering the tertiary tube settler/filter unit was being broken up, due to high velocities created by constrictions at several points in the inlet distribution system to the tube bundles. In January 1971, modifications were made in the settler compartment to minimize these constrictions. It was difficult, however, to determine the extent to which the modifications improved the performance of the tertiary system.

The tertiary filter unit was originally installed with a rotating surface wash system to break up and scour the surface of the bed during the backwash cycle. Repeated bearing failures in the washarms, as well as a continual loss of media during the surface wash operation, necessitated a design change. In April 1971, a fixed grid system was installed over the filter bed, replacing the rotary washarms. This change considerably reduced media loss and eliminated the previous maintenance problem.

CHEMICAL SLUDGE DECANT PUMP

The chemical sludge decant pump motor burned out on two occasions, as a result of freezing weather conditions in November 1970 and February 1971.

SURFACE WASH DIAPHRAGM VALVE

Freezing conditions in January 1971 also damaged the surface wash diaphragm valve, forcing suspension of routine tertiary operation until warmer weather.

OTHER MECHANICAL PROBLEMS

Water collection in the pneumatic control lines damaged the tertiary flowmeter pressure transducer in December 1970. While the transducer was being repaired at the factory, a manometer was used to monitor the tertiary flow rate. A refrigeration type air dryer was installed to desiccate air for the plant pneumatic control systems.

On two occasions the plant flowmeter jammed. In both instances the unit was removed and inspected. The causes for jamming could not be determined, however.

SECTION VIII
DISCUSSION

SECONDARY TREATMENT SYSTEM

MICROBIOLOGY

So long as the D.O. remained above 0.5 mg/l, the relatively long sludge ages at which the secondary system operated tended to produce a rather diverse population of organisms, ranging from bacteria to rotifers.

During the latter phases of the program, when the D.O. hovered between 0.1 to 0.2 mg/l, the higher forms all but disappeared. Surprisingly, few filamentous organisms were observed at these low D.O. concentrations.

SLUDGE YIELD

The quantity and apparent resistance to biological breakdown of the influent suspended solids had a significant affect on the gross sludge yield in the secondary treatment system. The gross sludge yield is defined as the total pounds of solids (including non-metabolized influent suspended solids as well as those biologically synthesized) lost from the secondary system per pound of influent BOD_5 removed.

Figure 42 is a plot of gross sludge yield as a function of sludge age.

Each data point represents the 5-day weekly average of the solids lost and influent BOD_5 removed over a temperature range of 8 to 22 degrees C. An average BOD_5 removal of 98 percent was used in calculating the gross sludge yield, based on the average tertiary effluent soluble BOD_5 of 6 mg/l measured over the demonstration period.

The considerable scatter in the data is attributed to several factors:

1. The majority of the BOD_5 and TSS loading resulted from the industrial wastes discharged by the Hervin Company. The TSS and BOD_5 varied considerably from day to day, depending upon the product being processed and production rate.

2. Daily flows were measured at the plant effluent pump station. Due to the low influent flows and surge capacity of the aeration-surge basin, the daily influent and effluent flows were not necessarily equivalent. Thus, the actual daily BOD_5 and TSS loads could not be accurately determined.

3. BOD_5 analyses were normally made two or three days per week, while the TSS was determined daily. Estimates of the influent BOD_5 were made for those days when BOD_5 data were not available from plots of the influent TSS versus BOD_5.

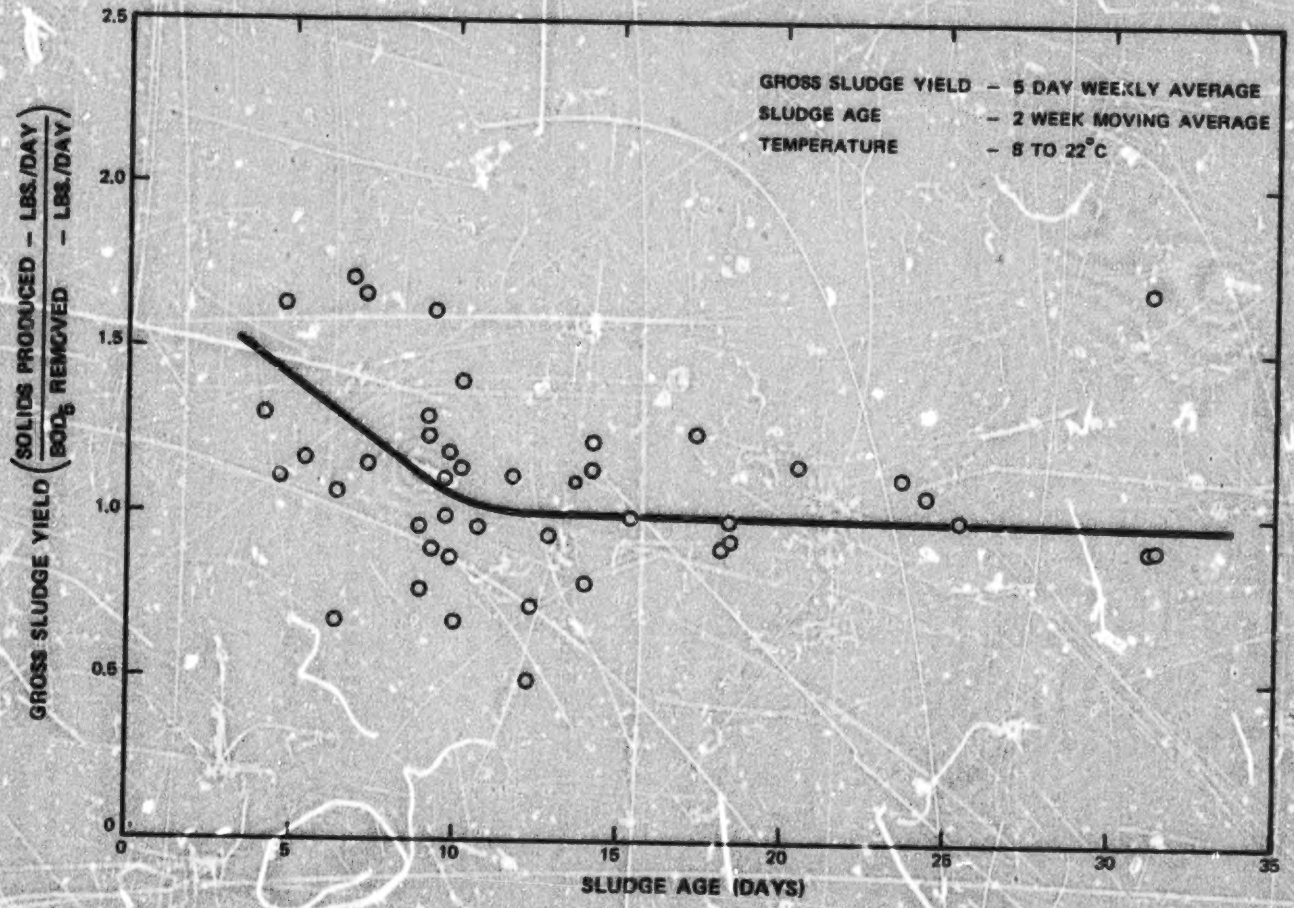


FIGURE 42

GROSS SLUDGE YIELD VERSUS SLUDGE AGE

4. The temperature in the aeration-surge basin varied from 8 to 22 degrees C, averaging 15.5 degrees C. The data on Figure 42 were initially plotted for three temperature ranges, but correlation of the data was not significantly improved. While temperature undoubtedly had some affect on solids yield, the variability introduced by the first three factors appeared to mask any changes due to this parameter.

The gross sludge yield declined from about 1.5 pounds per pound of BOD_5 removed at a sludge age of 4 days, to 1.0 pounds per pound of BOD_5 removed at a sludge age of 10 to 11 days. The gross sludge yield remained essentially constant from a sludge age of 11 to 31 days.

The influent suspended solids were suspected to be responsible for the larger than anticipated solids production. An attempt was made to calculate the biologically synthesized solids fraction of the gross sludge yield, by subtracting out the non-metabolized influent suspended solids. Assuming the rate of degradation to be dependent upon sludge age, the following equation was developed to estimate the non-metabolized fraction of the influent TSS at a given sludge age and temperature condition:

$$\underbrace{\frac{WAS + TSS_{(S)} + \Delta MLSS}{BOD_{(R)}}}_{\text{gross sludge yield}} - \frac{f \times TSS_{(I)}}{BOD_{(R)}} = \underbrace{NSY}_{\text{net sludge yield}} \Bigg|_{T_s,\, T}$$

Where:

WAS - waste activated sludge (lbs/day)

$TSS_{(S)}$ - secondary effluent total suspended solids (lbs/day)

$TSS_{(I)}$ - influent total suspended solids (lbs/day)

$\Delta MLSS$ - change in mixed liquor suspended solids (lbs/day)

$BOD_{(R)}$ - influent BOD_5 removed (lbs/day)

f - non-metabolized fraction of $TSS_{(I)}$

NSY - net sludge yield (i.e., total solids biologically synthesized less endogenous respiration loss)

T_s - sludge age (days)

T - temperature (degrees C)

This relationship takes the form of the straight line equation:

$$y = fx + b$$

Where: y = gross sludge yield

$$x = \frac{TSS_{(I)}}{BOD_{(R)}}$$

b = net sludge yield (NSY)

f = slope (non-metabolized fraction of $TSS_{(I)}$)

The gross sludge yield was plotted against $TSS_{(I)}/BOD_{(R)}$ to estimate the non-metabolized fraction, f, of the influent TSS for various sludge age conditions. The limited amount of data at any given sludge age necessitated using values obtained over a temperature range of 8-22 degrees C. Figure 43 is an example of one of these plots for a sludge age of 12-16 days.

The non-metabolized fraction, f, determined from these plots is shown as a function of sludge age in Figure 44. It appears that essentially no degradation of the influent TSS occurred at sludge ages less than 5 days and that only 20 to 22 percent of the influent TSS was ultimately biologically metabolized at sludge ages greater than about 10 to 12 days.

The impact of the influent TSS on such operational parameters as the F/M and activated sludge wasting rates can be significant, particularly in activated sludge systems where primary treatment is not provided. At influent conditions where the TSS and BOD_5 are approximately equal, the apparent F/M based on MLSS may be lower by a factor of two or more than the F/M considering only the biologically synthesized solids. The MLSS level necessary to provide the concentration of active biological solids desired must allow for the non-degraded influent TSS. Consideration must also be given to the influent TSS loading in estimating sludge wasting rates and sizing of sludge handling and storage facilities.

The degradability factor, f, determined above was applied to the influent TSS to calculate the net sludge yield for various sludge ages. The results are shown on Figure 45. The net sludge yield decreased from about 0.4 to 0.25 pounds per pound of BOD_5 removed between sludge ages of 5 and 12 days. Beyond a 12-day sludge age, the net sludge yield decreased at a much slower rate down to a minimum of about 0.17 pounds per pound of BOD_5 removed at about a 30 day sludge age.

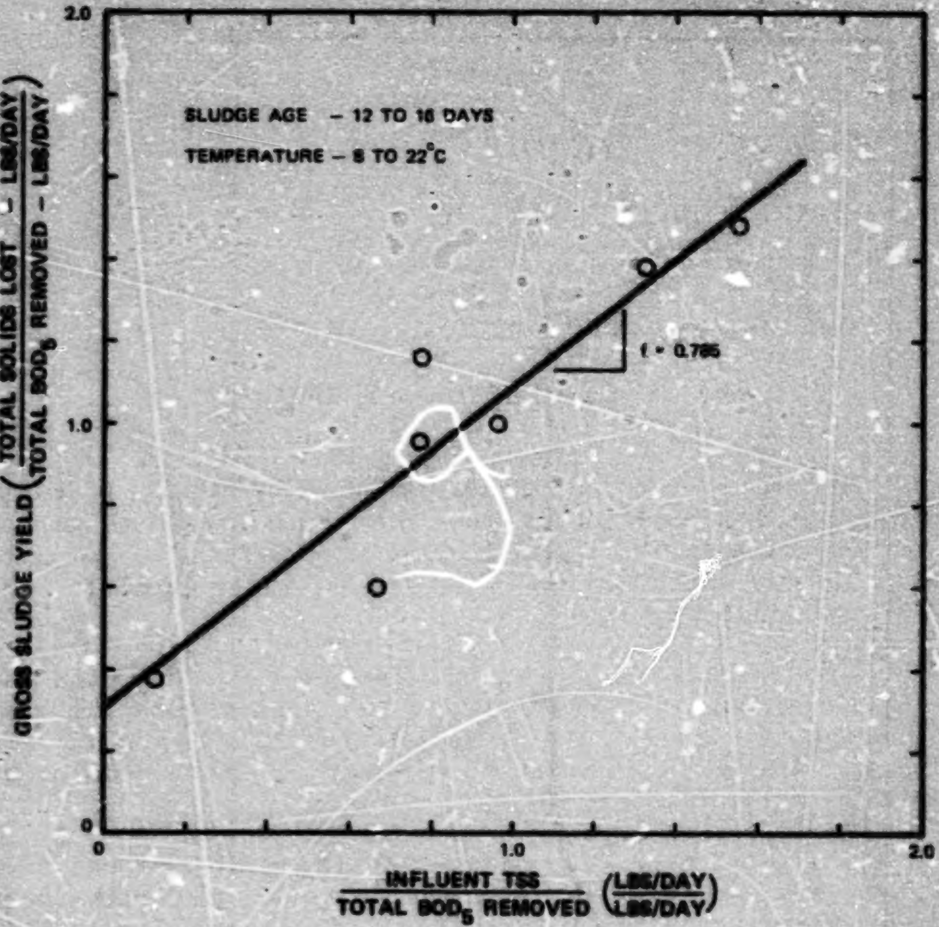


FIGURE 43

GROSS SLUDGE YIELD VERSUS
INFLUENT TSS/TOTAL BOD_5 REMOVED

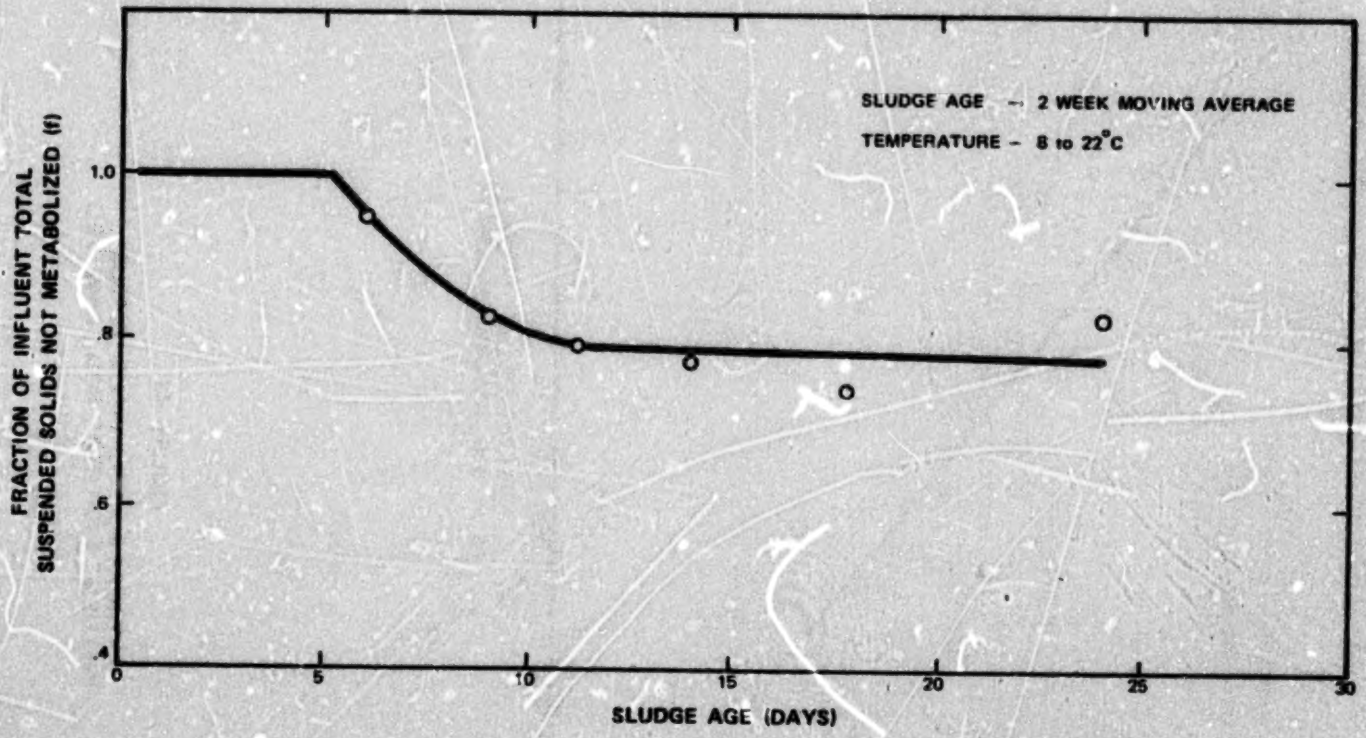


FIGURE 44

INFLUENT TOTAL SUSPENDED SOLIDS DEGRADATION VERSUS SLUDGE AGE

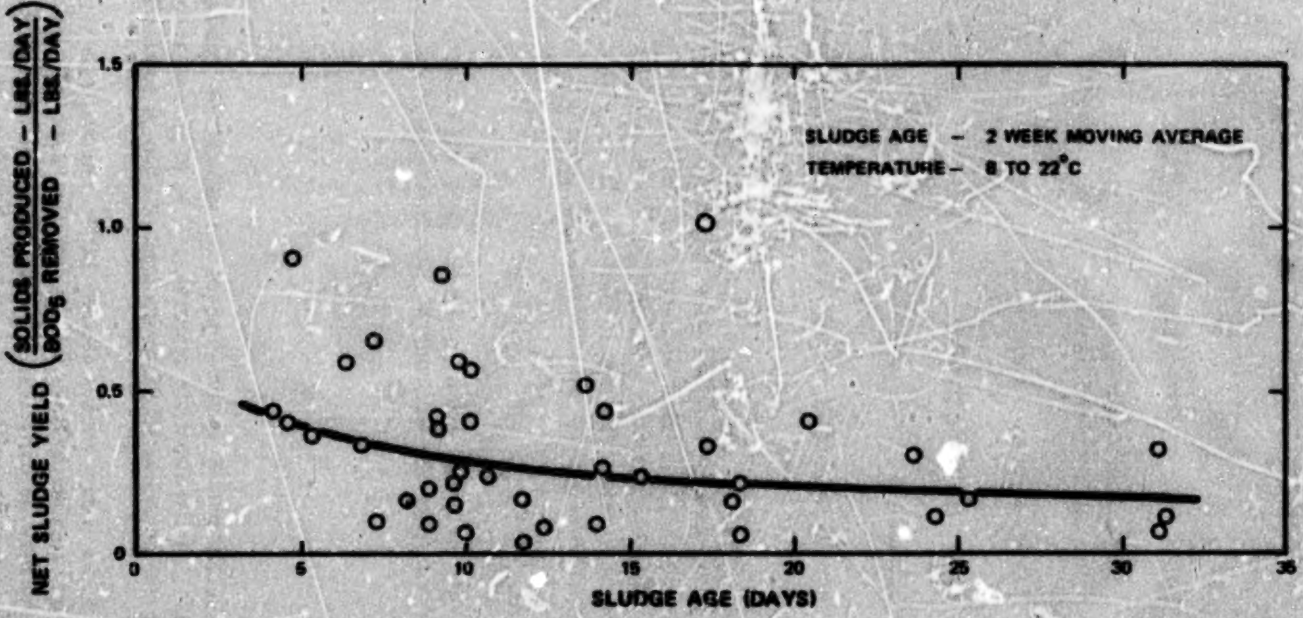


FIGURE 45

NET SLUDGE YIELD VERSUS SLUDGE AGE

The sludge yield and endogenous coefficients were evaluated using the general form of the equation developed by McCarty and Brodersen [10] and discussed in Section III:

$$\frac{A}{F} = a - \frac{0.8a}{1 + \frac{1}{bT_s}}$$

Where: $\frac{A}{F}$ = net biological solids synthesized per pound of BOD removed

a = sludge yield coefficient

b = endogenous respiration coefficient

T_s = sludge age

The factor, 0.8 a, assumes 20 percent of the biologically synthesized solids are relatively resistant to biological oxidation [10].

For sludge ages between 5 and 30 days and a temperature range of 8 to 22 degrees C, the sludge yield coefficient, a, was calculated to be 0.60 and the endogenous coefficient, b, to be 0.15. These values are in general agreement with values of 0.65 and 0.18 determined by McCarty and Brodersen for domestic sewage. Sludge yield and endogenous constants of 0.70 and 0.1 were determined for a similar extended aeration system treating a combined industrial and domestic waste at Dallas, Oregon [37].

SUBSTRATE REMOVAL

Substrate removal in the extended aeration system was evaluated in terms of the following equation derived from the Michaelis-Menton relationship discussed in Section III:

$$R = ks$$

The substrate removal coefficient, k, is the slope of the straight line passing through the origin produced by a plot of the substrate removal rate, R, expressed as total pounds of BOD_5 removed per day per pound of MLSS versus mg/l of effluent soluble BOD_5.

Figure 46a is a plot of the 5-day weekly average BOD_5 removal rate versus the 5-day weekly average effluent soluble BOD_5 for all of the data obtained during the demonstration period. As may be expected, the field data show considerable scatter. To determine the substrate removal coefficient, a straight line from the origin was drawn through the mean coordinates of the data points.

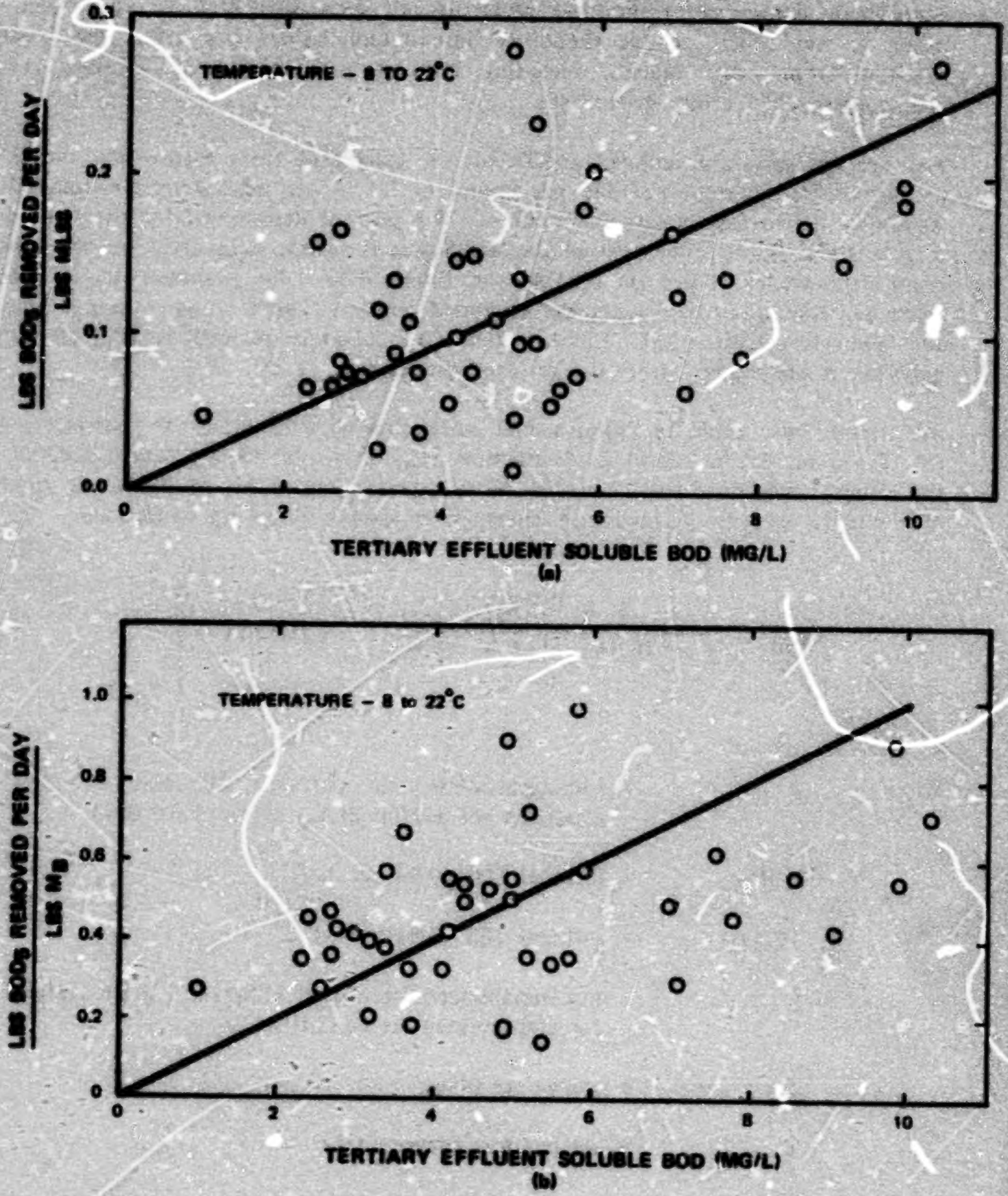


FIGURE 46

SUBSTRATE REMOVAL RATE VERSUS TERTIARY EFFLUENT SOLUBLE BOD_5

Figure 46a represents data for a temperature range of 8 to 22 degrees C. Initially, separate plots were made of the data in three temperature ranges. The substrate removal coefficients determined from these plots did not show significant variation from each other. It was concluded the variability due to temperature was within the variability resulting from other factors, including influent suspended solids loading and the characteristics of the industrial waste.

The average substrate removal coefficient, k, calculated from Figure 46a was 0.0235 pounds BOD_5 removed per day per pound of MLSS per mg/l soluble effluent BOD_5. Assuming the average volatile content of 73.5 percent determined for the WAS during August 1971 was representative of the MLSS over the entire demonstration, the substrate removal coefficient based on MLVSS was 0.032. This value is somewhat less than the value of 0.041 reported for another extended aeration plant treating combined industrial and domestic wastes [37]. The lower rate of substrate removal is attributed to the industrial waste received from the pet food processor.

An attempt was made to improve data correlation by calculating the substrate removal rate based on the biological solids fraction, M_B, of the MLSS. M_B was calculated from the influent suspended solids degradation rate data in Figure 44 and the sludge yield data in Figure 45 using the following relationship for various sludge age conditions:

$$M_B = \frac{y \times BOD_{(I)}}{y \times BOD_{(I)} + f \times TSS_{(I)}} \times [MLSS] \Bigg|_{T_S,\, T}$$

Where: y = biological solids yield coefficient (lbs solids produced per day/lbs BOD_5 removed per day)

$BOD_{(I)}$ = influent BOD_5 (lbs per day)

$TSS_{(I)}$ = influent TSS (lbs per day)

f = non-metabolized fraction of $TSS_{(I)}$ at a given sludge age and temperature condition

T_S = sludge age (days)

T = temperature (degrees C)

Figure 46b is a plot of the pounds of BOD_5 removed per pound of "biological" MLSS (M_B) versus effluent soluble BOD_5 concentration. Calculating substrate removal rate based on M_B rather than total MLSS did not markedly improve the data scatter, however. The substrate removal coefficient, k' based on M_B, was calculated from a straight line drawn through the mean coordinate of the data points and passing through the origin. The value of k' determined from Figure 46b was 0.098 pounds BOD_5 removed per day per pound M_B per mg/l effluent soluble BOD_5.

SLUDGE VOLUME INDEX

The sludge volume index was dependent to a large extent on the D.O. level in the aeration-surge basin. Figure 47 is a plot of the average weekly D.O. and SVI. As long as the D.O. remained above 0.5 to 1.0 mg/l, the SVI averaged about 80 and generally stayed below 100. Below 0.5 mg/l D.O., the SVI increased to values as high as 400.

Attempts were also made to relate SVI to F/M and sludge age. A correlation to either parameter was not apparent.

Although the SVI did not appear to have a significant effect on the performance of the secondary tube clarifier, high levels would undoubtedly reduce the degree of thickening obtainable in a conventional secondary clarifier and substantially influence activated sludge recycle and wasting rates.

SECONDARY CLARIFIER PERFORMANCE

The secondary effluent suspended solids concentration varied considerably, even after the air sparger modification was made to prevent entrapment of the solids and clogging of the tubes. Tube clarifier performance appeared to depend upon five variables:

1. MLSS concentration
2. Overflow rate
3. Temperature
4. pH
5. Extent of nitrification

Under a given set of conditions any of these parameters could become the controlling variable or considerably influence clarification efficiency.

MIXED LIQUOR SUSPENDED SOLIDS–The MLSS level appeared to be the most critical parameter. Figure 48 demonstrates the influence of MLSS on the secondary effluent suspended solids concentration for overflow rates of 0.58 - 0.75 gpm/ft^2 and 1.50 - 1.67 gpm/ft^2.

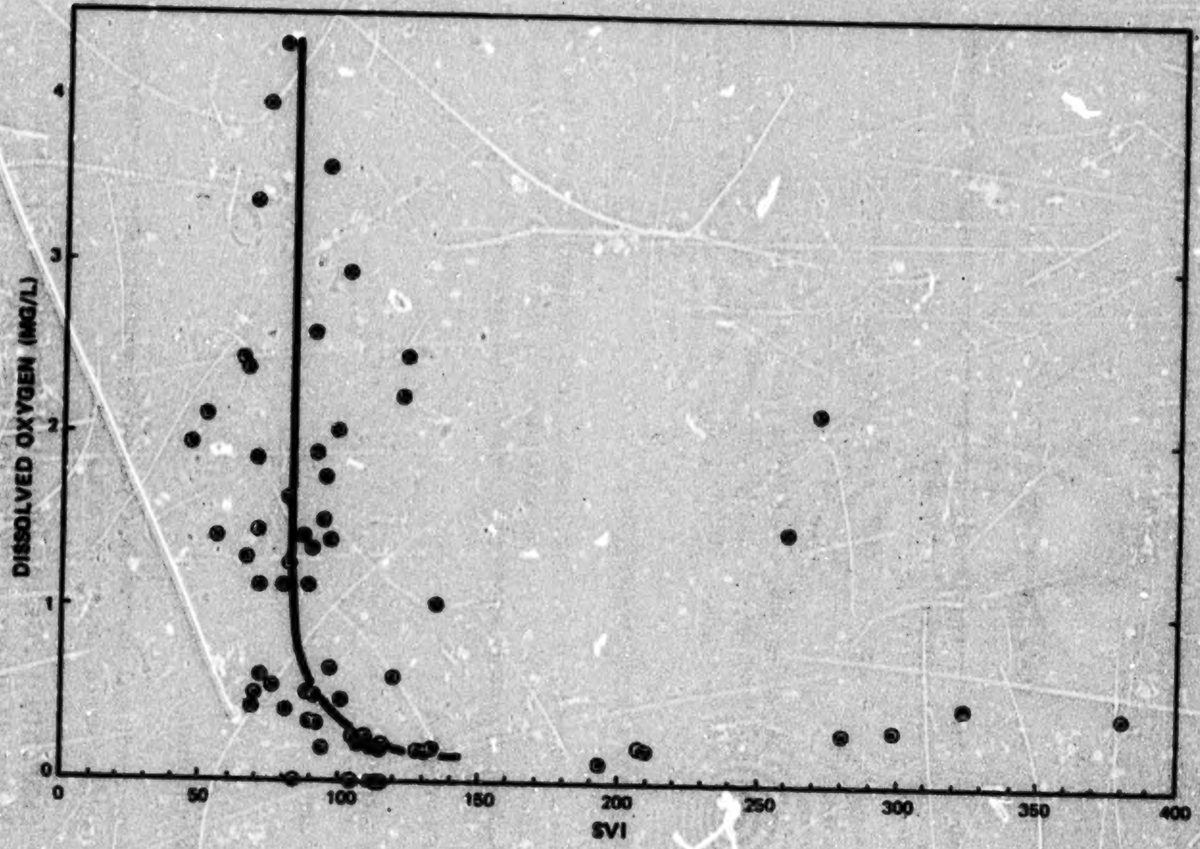


FIGURE 47

DISSOLVED OXYGEN VERSUS SVI

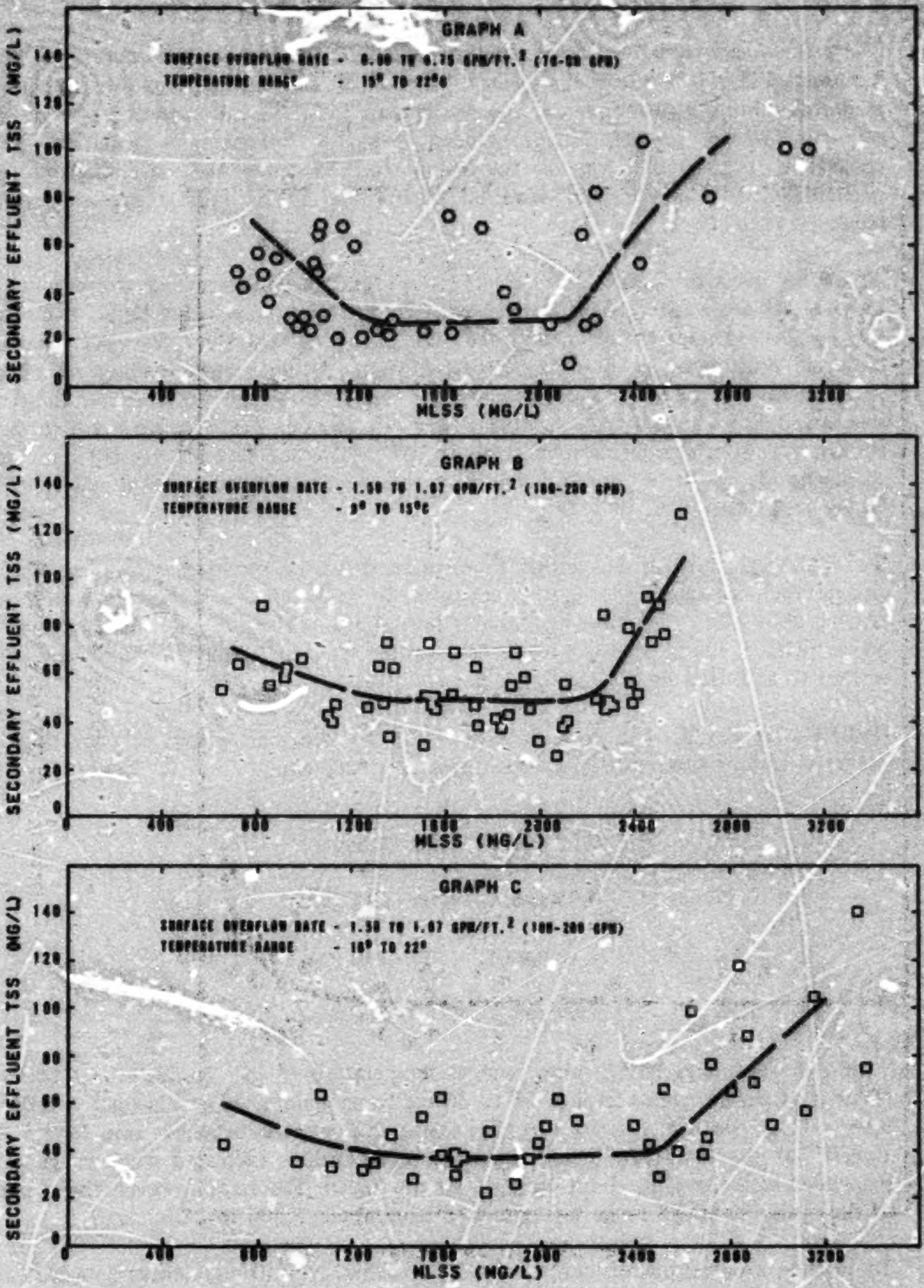


FIGURE 48

SECONDARY EFFLUENT SUSPENDED SOLIDS VERSUS MLSS

The effluent suspended solids concentration tended to decrease as increasing MLSS levels approached 1,000 - 1,200 mg/l. One or both of two mechanisms are thought to be responsible for this phenomenon: (1) An increase in MLSS concentration provides greater opportunity for collision of suspended particles, enhancing coagulation. (2) The coagulated solids collecting in the tubes form a filter bed, straining out smaller particulate. A minimum concentration of 1,000 - 1,200 mg/l is necessary to effect optimum removal.

At MLSS levels of 2,200 to 2,400, a critical condition was reached above which solids removal efficiencies sharply decreased. It is hypothesized that rising MLSS levels increase the volume of material settling in the tubes. The upward flow cross sectional area is reduced, resulting in an effective increase in the hydraulic surface loading. At a MLSS of 2,200 to 2,400 mg/l, a critical velocity condition is reached. Particles less than a given size and density do not have sufficient time to settle out before being swept through the tubes and into the collection pool above. The increased upward velocity may also be sufficient to produce a scouring condition, resuspending settled material and contributing to the solids carryover.

By modifying the clarifier design to increase the cross sectional area of the tubes or change their geometry, it may be possible to increase the MLSS range of optimum settling above the levels observed in this investigation. However, the constraint of MLSS on suspended solids removal must be considered a limitation in application of tube settlers to activated sludge clarification.

OVERFLOW RATE–The secondary tube clarifier operated within four overflow rate ranges at various times during the demonstration program:

1. 0.58 - 0.75 gpm/ft^2 (70 - 90 gpm)
2. 0.87 - 0.95 gpm/ft^2 (105 - 115 gpm)
3. 1.12 - 1.16 gpm/ft^2 (135 - 140 gpm)
4. 1.5 - 1.67 gpm/ft^2 (180 - 200 gpm)

Within the optimum MLSS range (and at temperature of 16 - 22 degrees C) a secondary effluent suspended solids level of 25 to 30 mg/l was maintained at a surface overflow rate 0.58 - 0.75 gpm/ft^2 (Figure 48). Increasing the surface overflow rate to 1.5 - 1.67 gpm/ft^2 at the same MLSS and temperature conditions increased the average effluent suspended solids level to about 40 mg/l. At the higher flow rate, however, the upper limit of the optimum MLSS range was extended from about 2,200 to 2,500 mg/l.

To provide a consistent effluent quality of less than 30 - 40 mg/l, these data indicate the surface overflow rate should be limited to 0.75 gpm/ft^2 or less at temperatures above 15 degrees C.

TEMPERATURE—A decrease in temperature appeared to result in increased solids carryover. Secondary effluent suspended solids are plotted versus MLSS for an overflow rate of 1.5 to 1.67 gpm/ft^2 and temperature ranges of 8 - 15 degrees C and 16 - 22 degrees C on Figure 48. Under optimum MLSS conditions, the secondary suspended solids averaged about 40 mg/l at water temperature of 16 - 22 degrees C. Reducing the temperature range to [illegible] - 15 degrees C raised the average secondary effluent suspended solids concentration to about 50 mg/l. The 10 mg/l increase in solids carryover is attributed to increased water viscosity at the lower temperatures.

pH AND NITRIFICATION—A correlation was observed between pH, the ammonium ion concentration and the secondary effluent suspended solids concentration. Figure 49 is a plot of pH versus secondary effluent concentration at overflow rates of 0.92 gpm per ft^2 and 1.12 gpm per ft^2. The ammonium concentration versus suspended solids content is shown on Figure 50 for all four overflow rates. The data presented in Figures 49 and 50 were limited to a MLSS range of 1,000 to 2,400 mg/l. Above and below these limits the affect of MLSS on the secondary effluent concentration predominated over the influence of the other variables.

The minimum secondary effluent suspended solids concentrations tended to occur between pH 6.9 to 7.2 and at ammonium concentrations around 5 mg/l. Both pH and ammonium concentrations were largely controlled by the extent to which the secondary treatment system nitrified. The level of ammonium in the secondary effluent corresponding to the minimum suspended solids concentrations occurred coincident with the optimum pH for settling. Lower concentrations of ammonium, however, were also recorded when the pH favored suspended solids removal. Thus, pH is suspected to exert a greater influence on coagulation and settling than ammonium ion. It may be that the ammonium concentrations corresponding to the lower secondary effluent suspended solids levels are just a reflection of the pH conditions resulting from nitrification.

The stability and the surface charge or zeta potential associated with colloidal particles is known to be affected by such potential determining ions as H^+, HCO_3^- and NH_4^+. Adsorption of these ions onto the particle surface can decrease (or increase) the apparent surface charge. Reduction in the surface charge and the resultant loss of stability enhances the tendency for coagulation.

Bacteria, organic matter resulting from the lysing and metabolism of cell tissue, and inorganic colloids, such as clay particles, tend to have a net negative surface charge in a slightly alkaline medium. Maintaining the pH within a specified range, together with the presence of ammonium ion (NH_4^+), would be expected to neutralize the negative surface charge and destabilize the suspended colloidal particles. Under these conditions, coagulation and flocculation of the MLSS should be enhanced and clarifier performance improved.

Further research is needed to more fully explore the effect of pH, the presence of specific ions, and oxidative state on coagulation and secondary clarification of activated sludge.

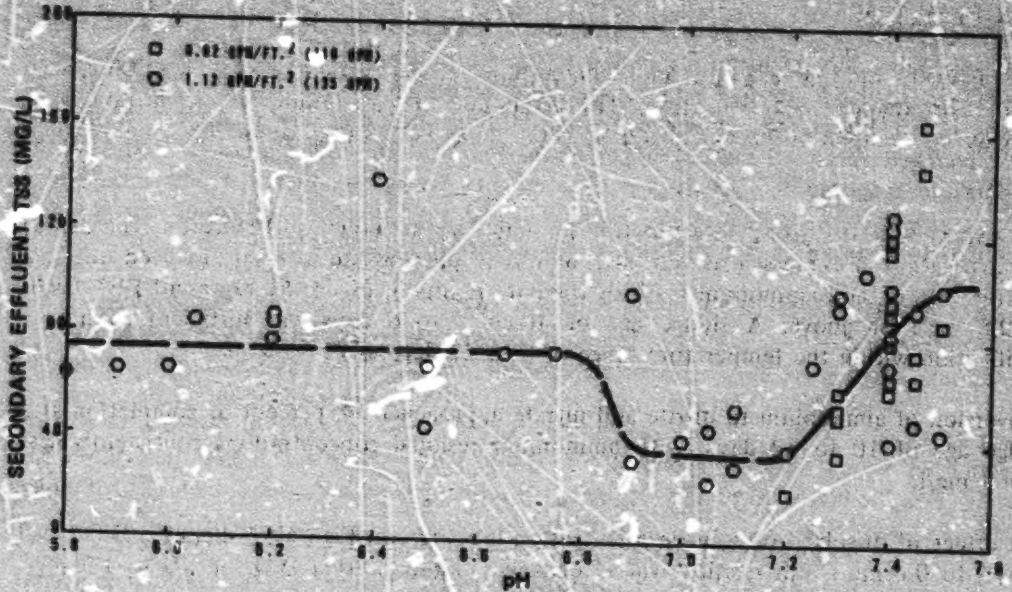


FIGURE 49

SECONDARY EFFLUENT SUSPENDED SOLIDS VERSUS pH

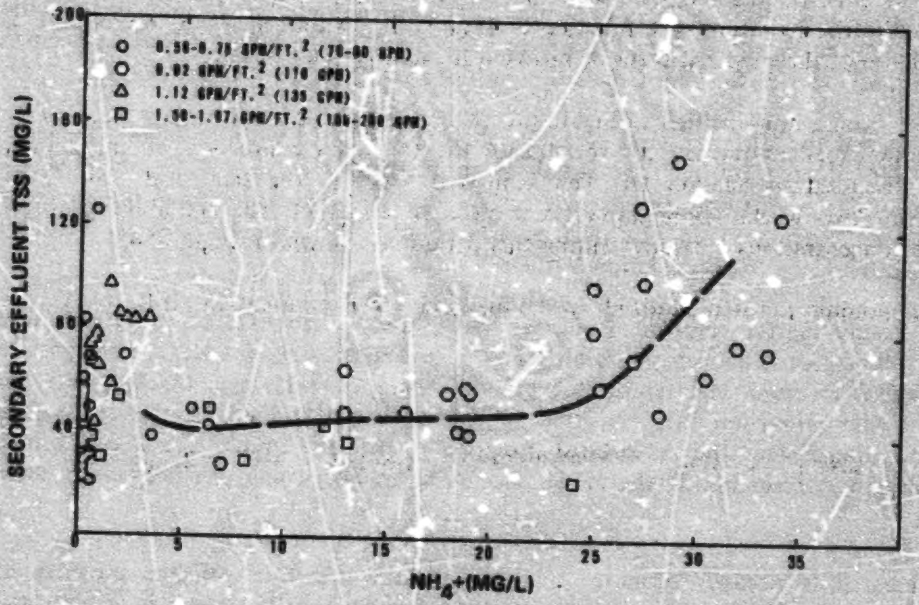


FIGURE 50

SECONDARY EFFLUENT SUSPENDED SOLIDS VERSUS AMMONIUM ION CONCENTRATION

NITRIFICATION

The relatively long sludge ages and temperatures in the secondary system produced an ideal environment for nitrification. The oxidation of ammonium ion to nitrate went virtually to completion, when adequate D.O. could be maintained in the aeration-surge basin.

The extent of nitrification as a function of sludge age is shown on Figure 51. The degree of nitrification is expressed as the ratio of nitrite plus nitrate to total dissolved nitrogen (nitrite, nitrate, and ammonium). All of the data points in Figure 51 represent D.O. levels of 0.8 mg/l or above. A sludge age of about 5 to 6 days was sufficient to induce nitrification within the temperature range of 8 to 22 degrees C.

Conversion of ammonium to nitrite and nitrate approached 98 percent of completion at a sludge age of 10 to 12 days, with ammonium residual concentrations consistently less than 1 mg/l.

The effect of dissolved oxygen on nitrification is shown on Figure 52. Below a D.O. level of 0.4 to 0.6 mg/l, nitrification was effectively inhibited. At a D.O. of 1.0 to 1.5 mg/l essentially complete (98 percent) oxidation was possible.

While warmer temperatures undoubtedly promote nitrification, particularly at the shorter sludge ages, the effect of temperature over a range of 8 to 22 degrees C appears to be less than the variability introduced in sampling and analysis in both Figures 51 and 52.

The effect of D.O. on the relative nitrite and nitrate concentrations was also examined. The ratio of nitrite to nitrate as a function of D.O. is shown on Figure 53. Above a D.O. of 0.6 to 0.8 mg/l the oxidized form of nitrogen is essentially all nitrate.

A transition from a non-nitrified to highly nitrified condition occurred in late June 1971, presenting an excellent opportunity to observe the two step conversion of ammonium nitrate as discussed in Section III. The industrial flows to the treatment plant were suspended for two weeks. Only domestic waste was processed. The ammonium, nitrite and nitrate concentrations measured during this period are shown on Figure 54.

Initially, ammonium ion was oxidized by the bacterium *nitrosomonas* to nitrite [11].

$$2NH_4^+ + 3O_2 \xrightarrow{\underline{nitrosomonas}} 2NO_2^- + 4H^+ + 2H_2O$$

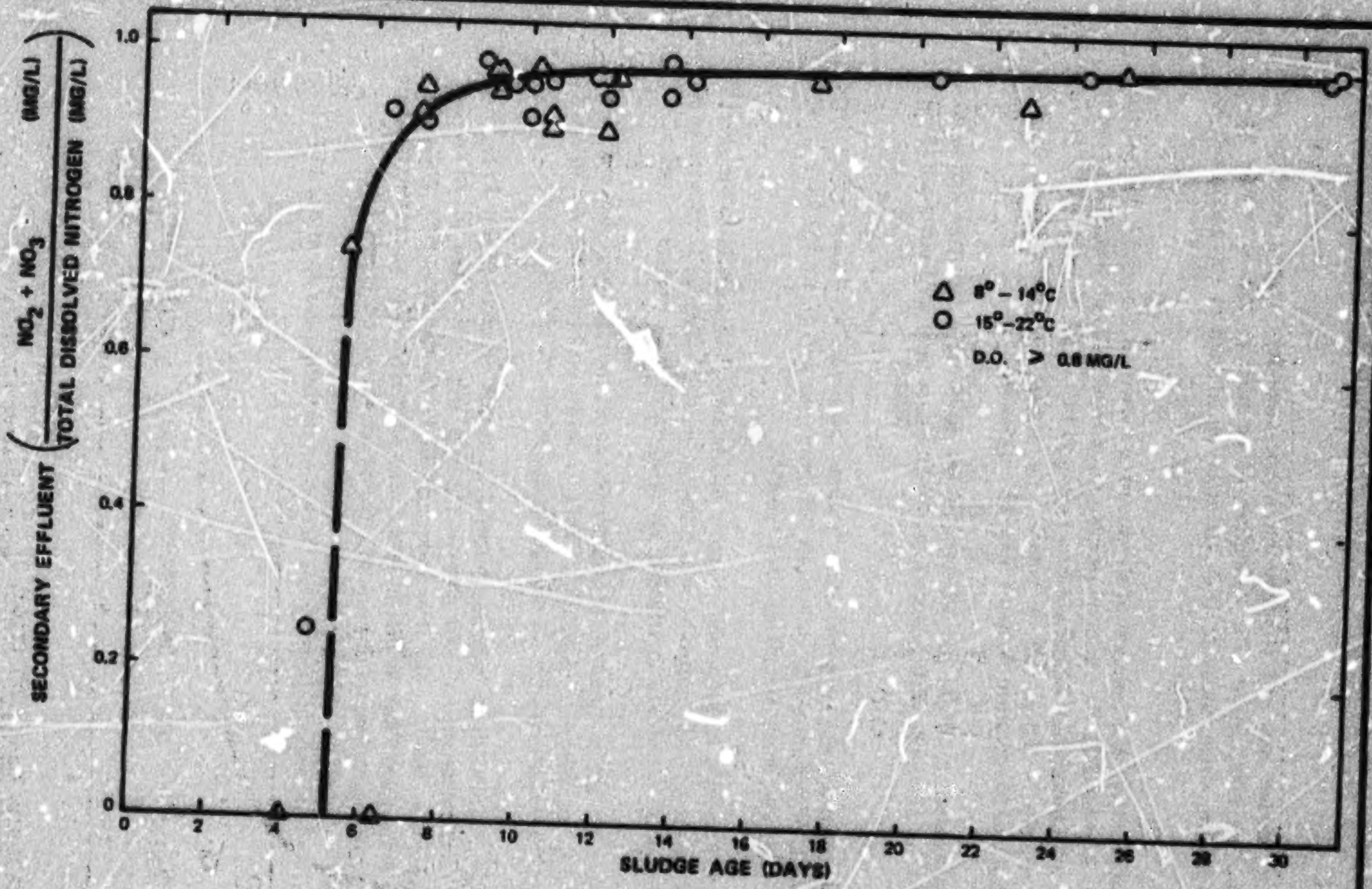


FIGURE 51

NITRIFICATION VERSUS SLUDGE AGE

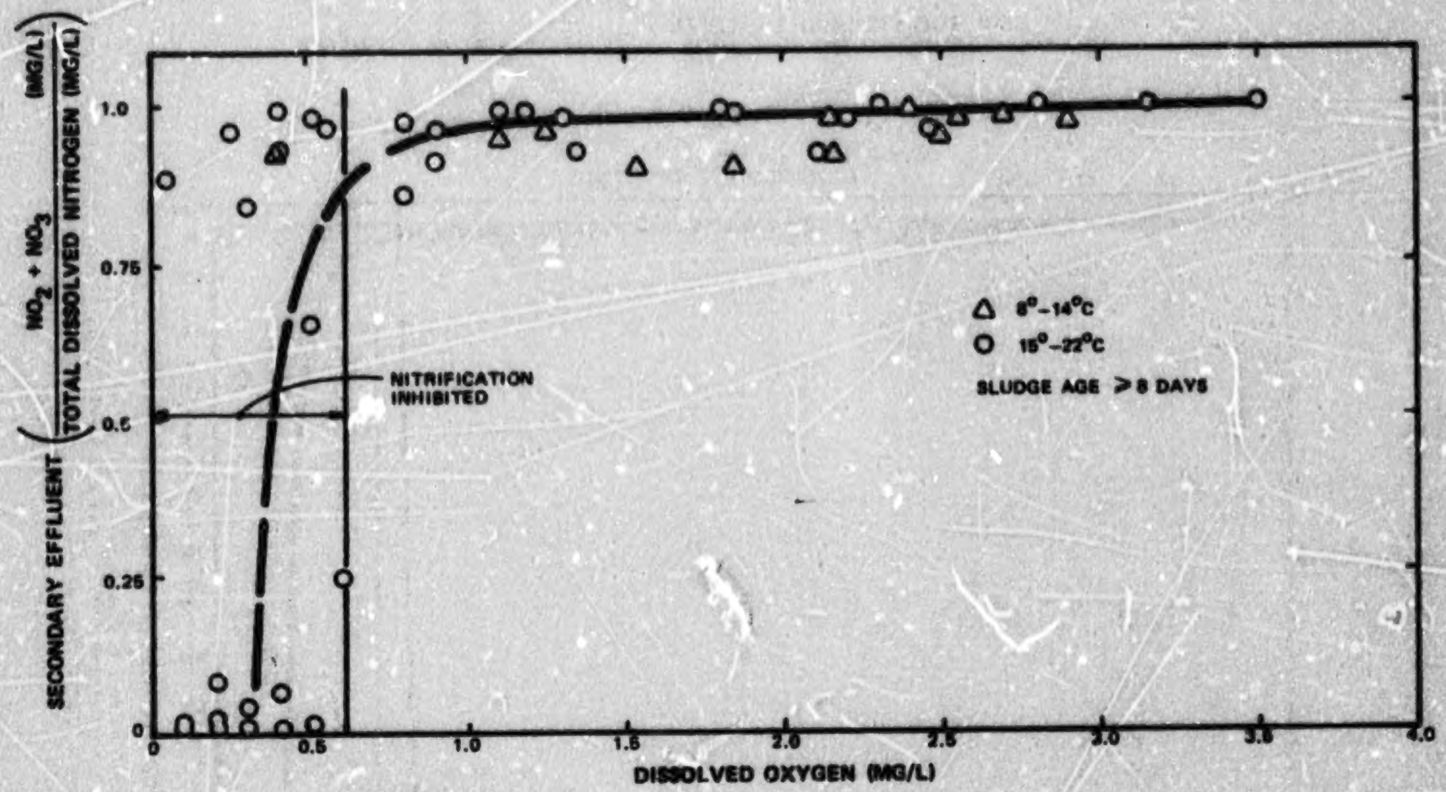


FIGURE 52

NITRIFICATION VERSUS AERATION–SURGE BASIN DISSOLVED OXYGEN

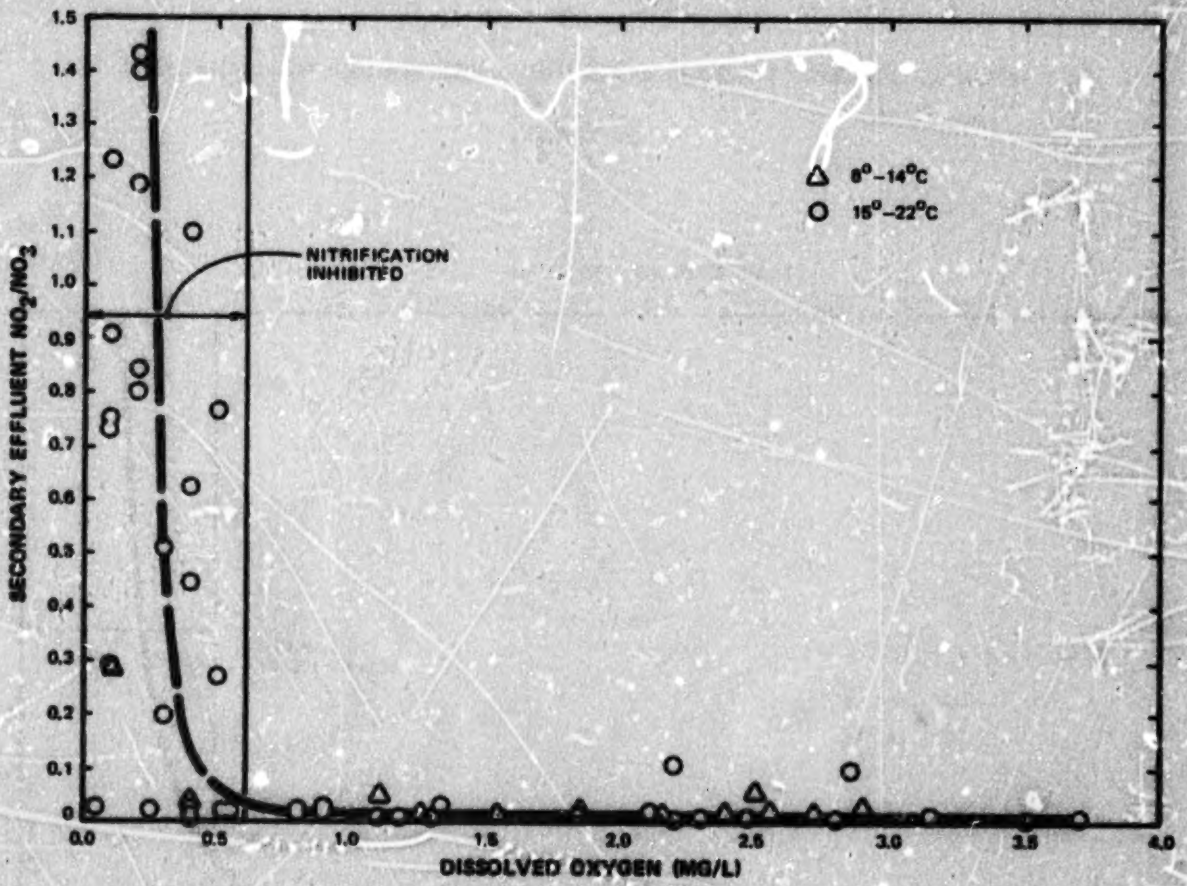


FIGURE 53

NO_2/NO_3 VERSUS AERATION–SURGE BASIN DISSOLVED OXYGEN

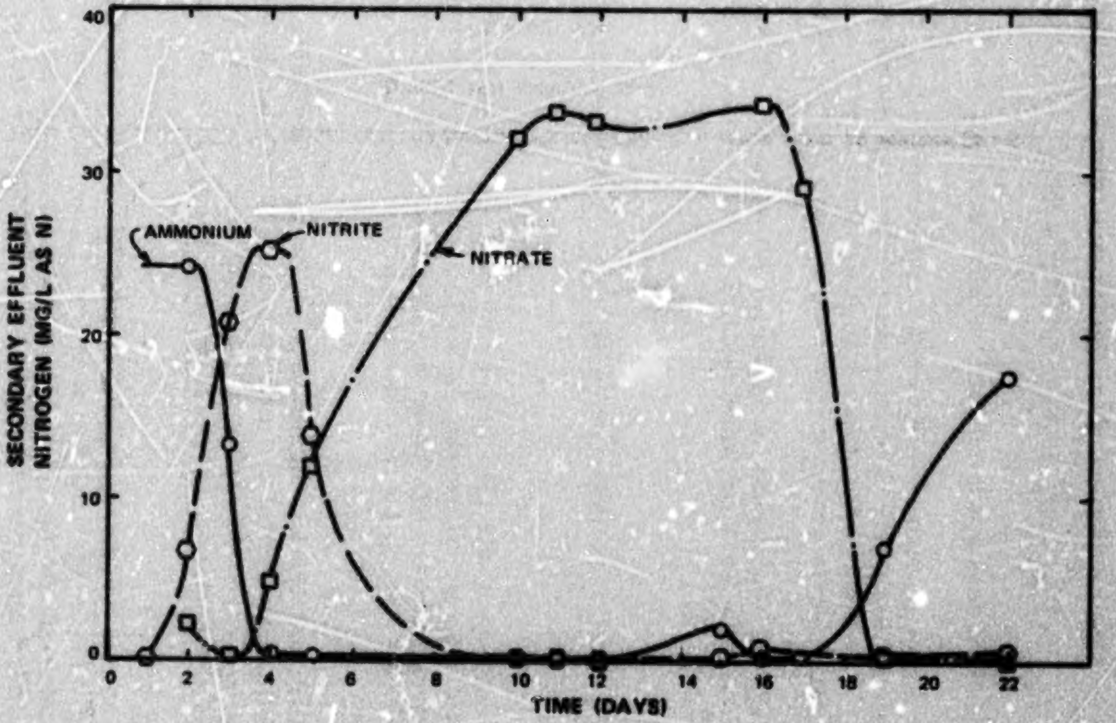


FIGURE 54

SEQUENCE OF NITRIFICATION REACTIONS

The concentration of ammonium decreased as the level of nitrite increased. About 48 hours after the start of nitrification, the bacterium *nitrobacter* began the conversion of nitrite to nitrate [11].

$$2NO_2^- + O_2 \xrightarrow{nitrobacter} 2NO_3^-$$

The concentration of nitrite declined, corresponding with the increase in nitrate.

When normal production operations were resumed by the Hervin Company, after the two week vacation period, the organic loading in the industrial flow exceeded the design aeration capacity. The nitrate concentration immediately dropped as bacteria scavenged for all available oxygen and the ammonium concentration began to steadily increase.

The bacterial oxidation of ammonium ion results in a net release of hydrogen ions and a corresponding reduction in alkalinity. According to the autotrophic nitrification reaction proposed by McCarty [12].

$$22NH_4^+ + 37O_2 + 4CO_2 + HCO_3^- \rightarrow C_5H_7NO_2 + 21NO_3^- + 20H_2O + 42H^+$$

6.98 mg of alkalinity (as $CaCO_3$) are required for each mg of ammonium ion (as N) oxidized to nitrate. Balanced on a weekly average basis, the actual observed reduction in total alkalinity during periods of nitrification generally agreed to within 10 percent of the theoretical requirement. Considering the sampling frequency, the degree of correlation is surprisingly good.

Several investigators have reported that nitrification is favored by an alkaline environment [13, 14, 15]. Nitrification occurred over a pH range of 5.7 to 7.45 and did not appear to be inhibited to any appreciable extent at the lower pH conditions. A 95 percent conversion was observed at a 12 day sludge age condition, when the pH was 5.75 and D.O. 2.5 mg/l.

The change in soluble nitrogen (ammonium, nitrite, and nitrate) concentration with sludge age was also briefly examined. The secondary effluent soluble nitrogen, expressed as percent of total influent nitrogen, is shown versus sludge age on Figure 55 for conditions when nitrification was both occurring and inhibited for lack of oxygen. As expected, the soluble nitrogen content increased with sludge age to between 12 and 13 days, as a result of endogenous respiration. Beyond a sludge age of 15 days there was a noticeable reduction in the dissolved nitrogen content. The reason for this apparent reduction in soluble nitrogen is not obvious. It is speculated that denitrification may occur at long sludge ages, even in the presence of dissolved oxygen.

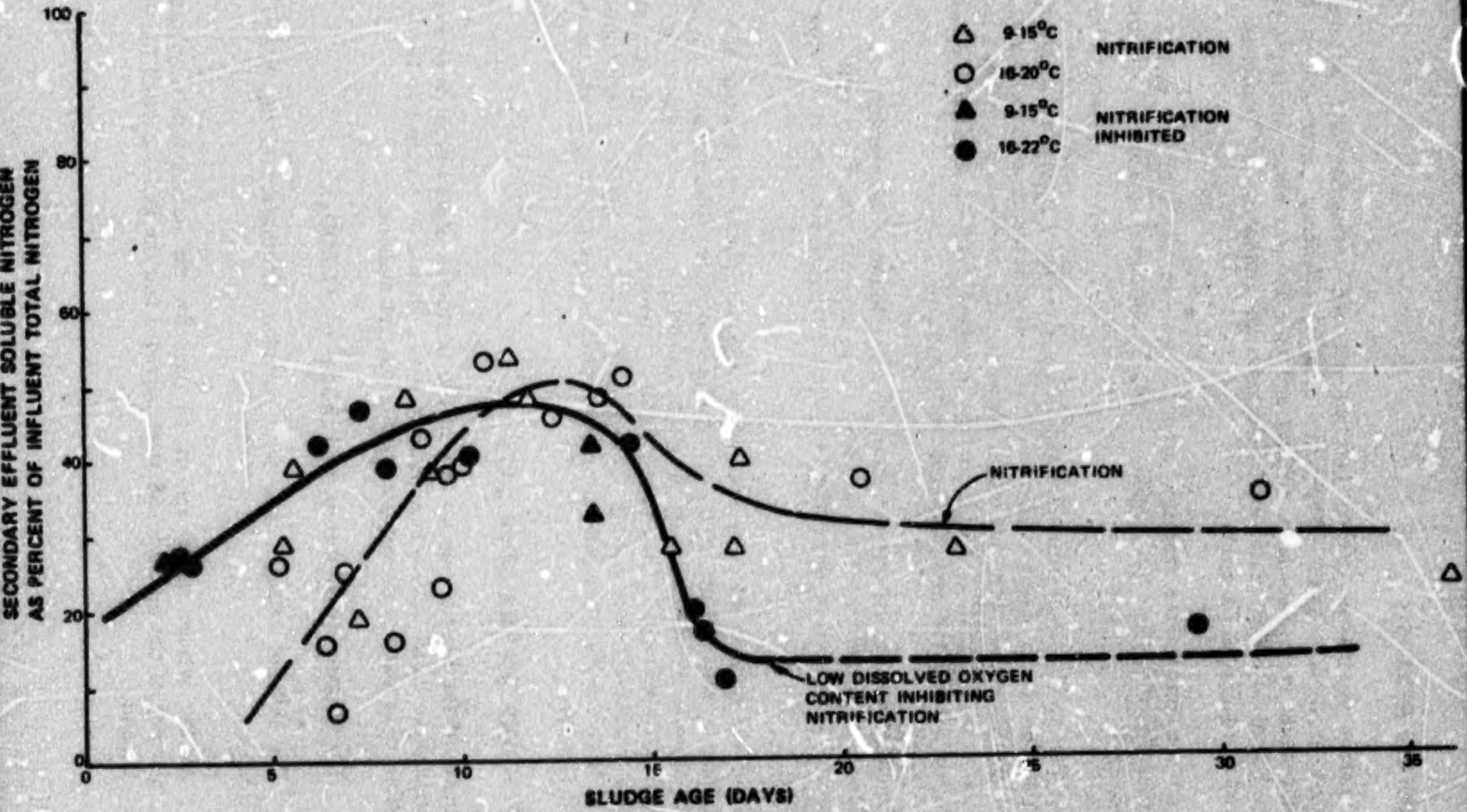


FIGURE 55

NITROGEN REMOVAL IN AERATION—SURGE BASIN

PHOSPHATE REMOVAL

An apparent correlation existed between pH and orthophosphate removal in the secondary treatment system. Phosphate removal efficiency, expressed in terms of orthophosphate in the secondary effluent and total influent phosphate is plotted versus pH on Figure 56. Each data point represents an average of the phosphate removals observed at a given pH condition.

Optimum orthophosphate removals occurred between a pH of 7.0 to 7.25. This is approximately the same pH region in which optimum secondary clarification was observed.

Whether the mechanism of orthophosphate removal is one of precipitation, adsorption, or other phenomenon is a matter of conjecture. Analyses for cation precipitants, such as calcium and aluminum, were not made. Therefore, it is not known whether a net loss of these elements occurred in the secondary treatment plant to an extent which would account for the observed orthophosphate reduction.

If orthophosphate removal was the result of precipitation with aluminum ion introduced by discharge of the tertiary filter backwash to the aeration-surge basin, optimum removal efficiencies, as observed in the tertiary system, should have occurred at or below a pH of 6.0 [25].

Because of the narrow pH range over which optimum phosphate removal occurred and the coincidence with the pH region of optimum suspended solid removal, it is suspected that the removal mechanism may be one of adsorption of orthophosphate on colloidal particulate. Orthophosphate may be a potential determining ion affecting the stability of suspended particulate. Considering the apparent effect of pH on coagulation of the MLSS, it is conceivable that optimum orthophosphate adsorption could also occur in this same pH regime.

AERATOR OXYGEN TRANSFER RATE

An estimate of the oxygen transfer rate of the floating aerators was made from the plot of aeration-surge basin D.O. versus influent BOD_5 shown on Figure 57. The following assumptions were made in the oxygen transfer rate calculations:

1. The ultimate BOD_L was exerted ($BOD_L = 1.47\ BOD_5$).
2. The influent nitrogen underwent complete nitrification (1 mg N = 3.84 mg O_2).
3. At the average sludge age of 13 days for the demonstration program, 20 percent of the influent TSS were degraded and the BOD_5 exerted the ultimate demand.

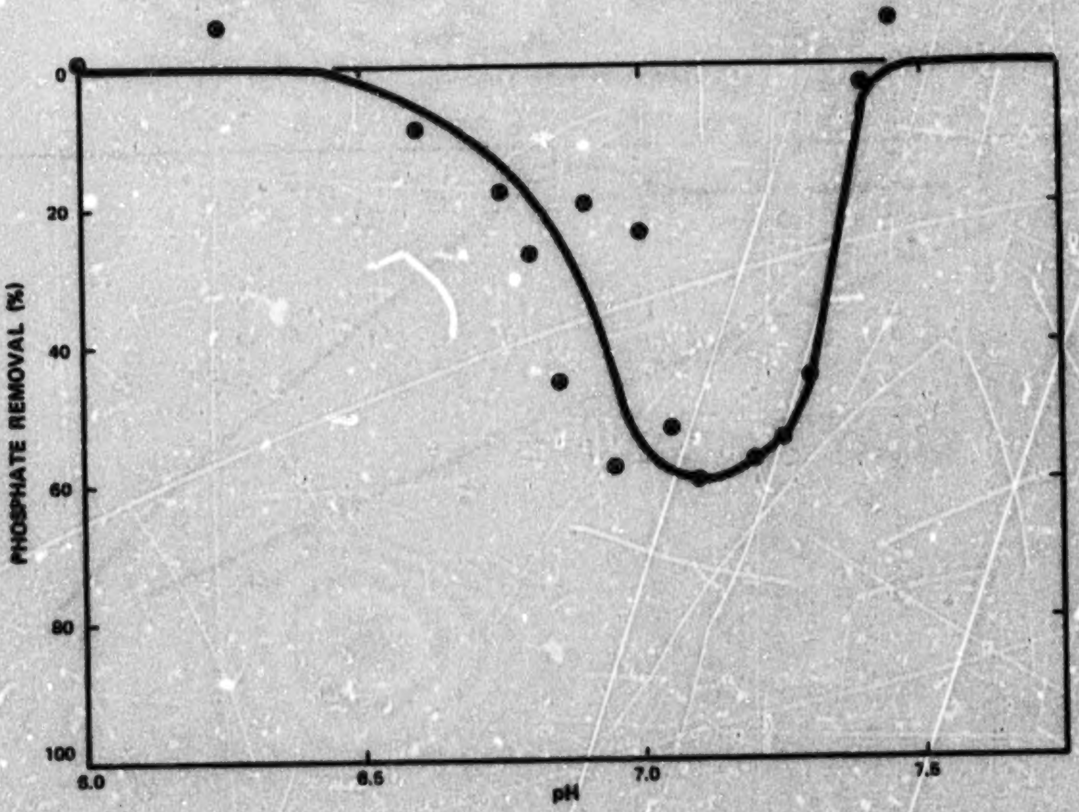


FIGURE 56

SECONDARY TREATMENT SYSTEM PHOSPHATE REMOVAL VERSUS pH

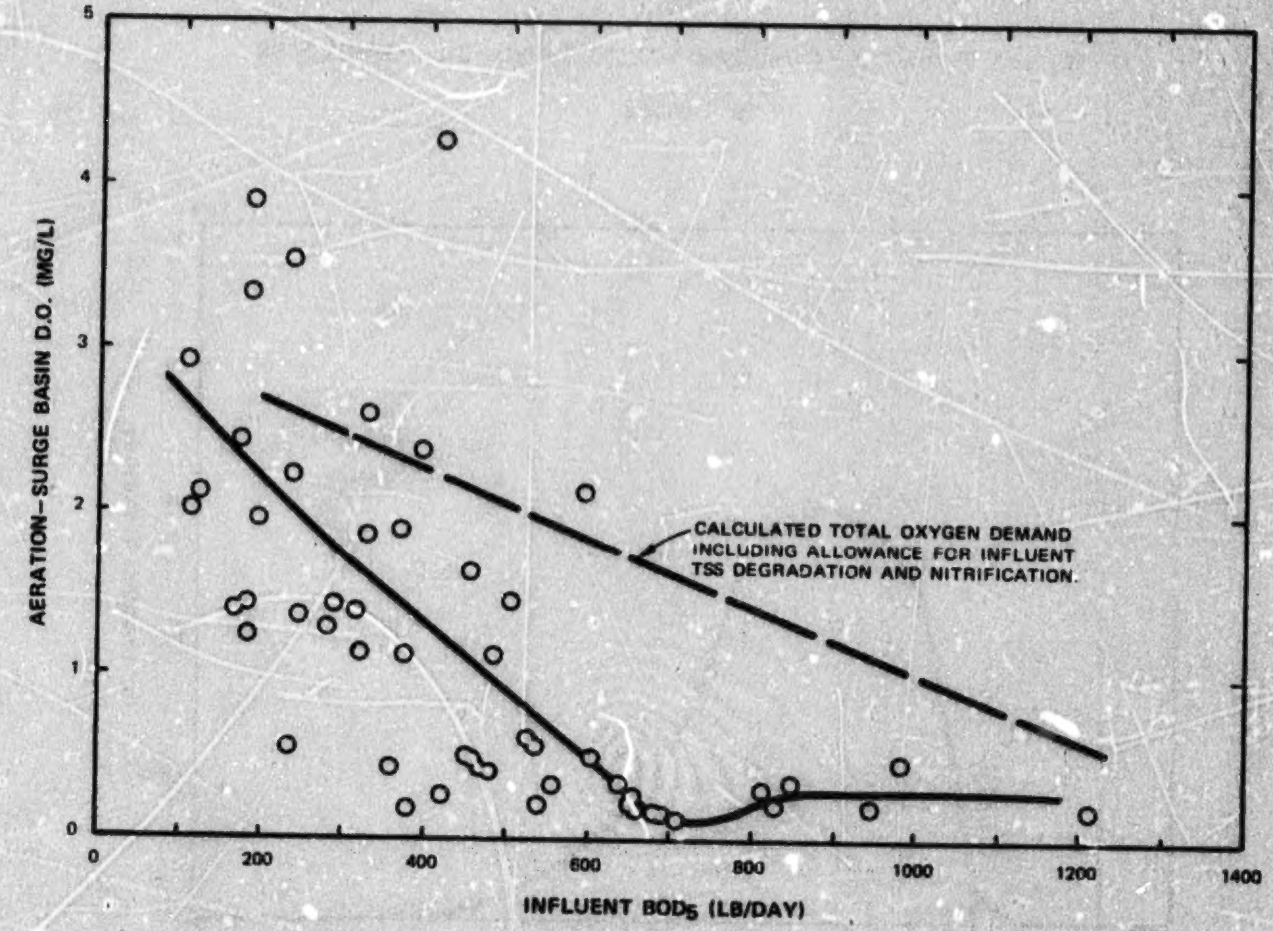


FIGURE 57

AERATION–SURGE BASIN D.O. VERSUS INFLUENT BOD$_5$

The influent BOD_5 and suspended solids concentrations were approximately equal, when averaged over the demonstration period. The influent total nitrogen averaged 10 percent of the influent BOD_5. Based on the three assumptions above, the total average oxygen requirement was calculated to be equivalent to 2.15 times the influent BOD_5 as follows:

BOD_L = 1.47 mg/l O_2/mg/l BOD_5

TSS = 1.47 x 0.2 mg/l O_2/mg/l BOD_5

N = 3.84 x 0.1 mg/l O_2/mg/l BOD_5

Total O_2Demand = 2.15 mg/l O_2/mg/l Influent BOD_5

At the average temperature condition of 15.5 degrees C and a D.O. level of 1.0 mg/l in the aeration-surge basin, the field transfer rate for the two 15 hp aerators was calculated to be 1.43 lbs O_2/hp-hr (nameplate hp basis). Assuming an α of 0.85 (ratio of oxygen transfer rate into waste to oxygen transfer rate in clean water) and β of 0.9 (ratio of saturation D.O. in waste to saturation D.O. in clean water) the clean water transfer rate of the aerators was calculated to be 2.13 lbs O_2/hp-hr. This transfer rate is somewhat below that expected for draft tube type aerators.

The estimate of aerator oxygen transfer rate must be somewhat qualified, however. The assumptions made in calculating the total oxygen demand allow for the maximum oxygen utilization that could reasonably be expected and may be on the high side. The data on Figure 57 used in the calculations represent a temperature range of 8 to 22 degrees C. D.O. measurements were made only during the day, while the BOD_5 analyses were made from samples composited over a 24-hour period and represent the average BOD_5 applied to the plant. The diurnal variation in BOD_5 loading was not measured. Peak organic loading conditions likely occurred during the daytime hours when the D.O. measurements were made.

TERTIARY TREATMENT SYSTEM

TERTIARY SYSTEM PERFORMANCE

The cycle time of the tertiary unit was influenced by four factors:

1. Premature carryover of solids from the tertiary tube settler on to the filter.
2. Secondary effluent suspended solids concentration.
3. Alum dosage.
4. Excessive polyelectrolyte addition.

SOLIDS CARRYOVER FROM THE TERTIARY TUBE SETTLER–The tertiary tube settler did not have sufficient solids storage capacity to retain all of the settleable chemical floc before entrapment of finer particulate by the filter media required backwashing of the bed. As a result, large quantities of chemical floc were prematurely carried over onto the filter bed, causing a rapid rise in head loss and substantial reduction in filter cycle time.

Figure 58 illustrates the effect of the solids carryover on filter performance. The suspended solids content of the effluent from the tertiary tube settler, filter head loss and tertiary effluent turbidity and suspended solids were monitored for one filter run. The filtration rate was 3.8 gpm/sq ft. The secondary effluent contained a TSS of 65 mg/l, to which an alum dosage of 182 mg/l was applied for coagulation. No polyelectrolyte was added either presettler or prefilter.

At about 1-1/2 hours into the run, solids began to carryover from the settler onto the filter. An increase in the rate of filter head loss buildup was observed at about 2-1/2 hours. At 3 hours, 56 minutes after start of the run, the head loss reached 8.3 feet, the preset value initiating the backwash cycle and terminating the run.

A total volume of 44,840 gal was filtered during the run. Had the storage volume in the tube settler been adequate to prevent carryover, extrapolation of the head loss buildup during the first 2 hours of the run indicates the filter cycle could have been extended from 4 to 16-1/2 hours.

The existing tube settler was designed for a surface overflow rate of 143 gal per day per sq ft. From the data in Figure 58, the tube settler surface area should be increased by a factor of 4, reducing the overflow rate to about 35 gal per day per sq ft. The volume required for backwash was 9,600 gallons. At a flow rate of 190 gpm and 4 hour filter cycle, 21 percent of the tertiary throughput was required for backwash. Optimizing the size of the tertiary tube settler to provide more solids storage volume would reduce the backwash requirement to about 5 percent of the throughput.

INFLUENCE OF SECONDARY TSS ON CYCLE TIME–Secondary TSS versus the volume filtered per cycle without the addition of chemical is shown on Figure 59. The throughput decreased exponentially at TSS concentrations above 60 to 70 mg/l and increased linearly at TSS below 60 mg/l. It is suspected that at the higher loadings, the solids collected on the surface of the filter preventing "indepth" filtration throughout the bed. Figure 59 illustrates the importance of providing efficient secondary clarification when chemical coagulation and sedimentation are not provided prior to filtration. At a secondary effluent TSS of 20-30 mg/l, the throughput across the 49.5 sq ft filter bed approached 500,000 gal per cycle.

Increasing alum dosages reduced the influence of the secondary effluent TSS on the length of filter run. Figure 60 is a plot of the volume filtered per cycle versus secondary effluent TSS for five alum feed conditions. Above an alum dosage of 183 to 190 mg/l, the influence of the secondary effluent TSS to the tertiary system was minimal.

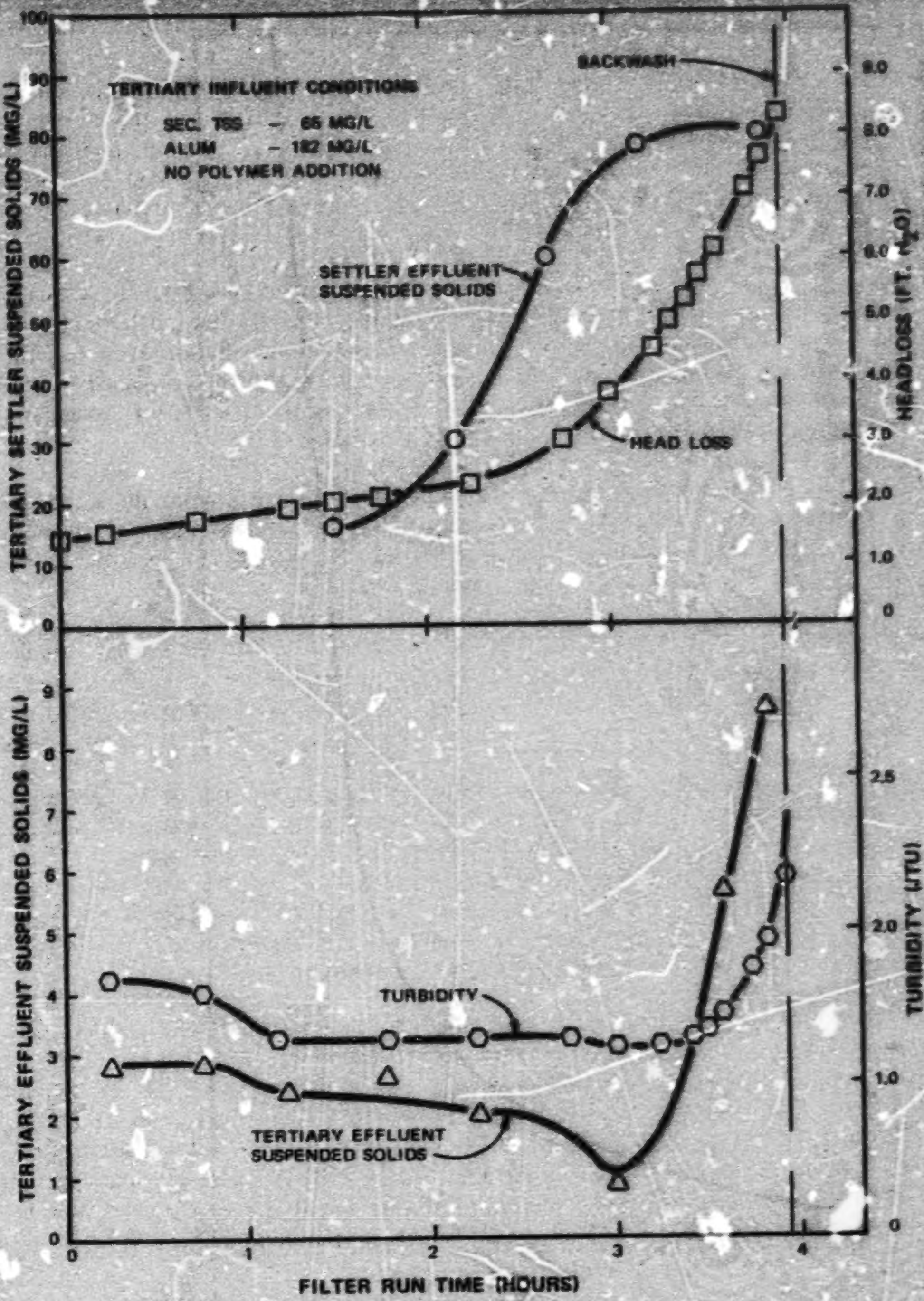


FIGURE 58

SUSPENDED SOLIDS, HEAD LOSS AND TURBIDITY VERSUS FILTER RUN TIME

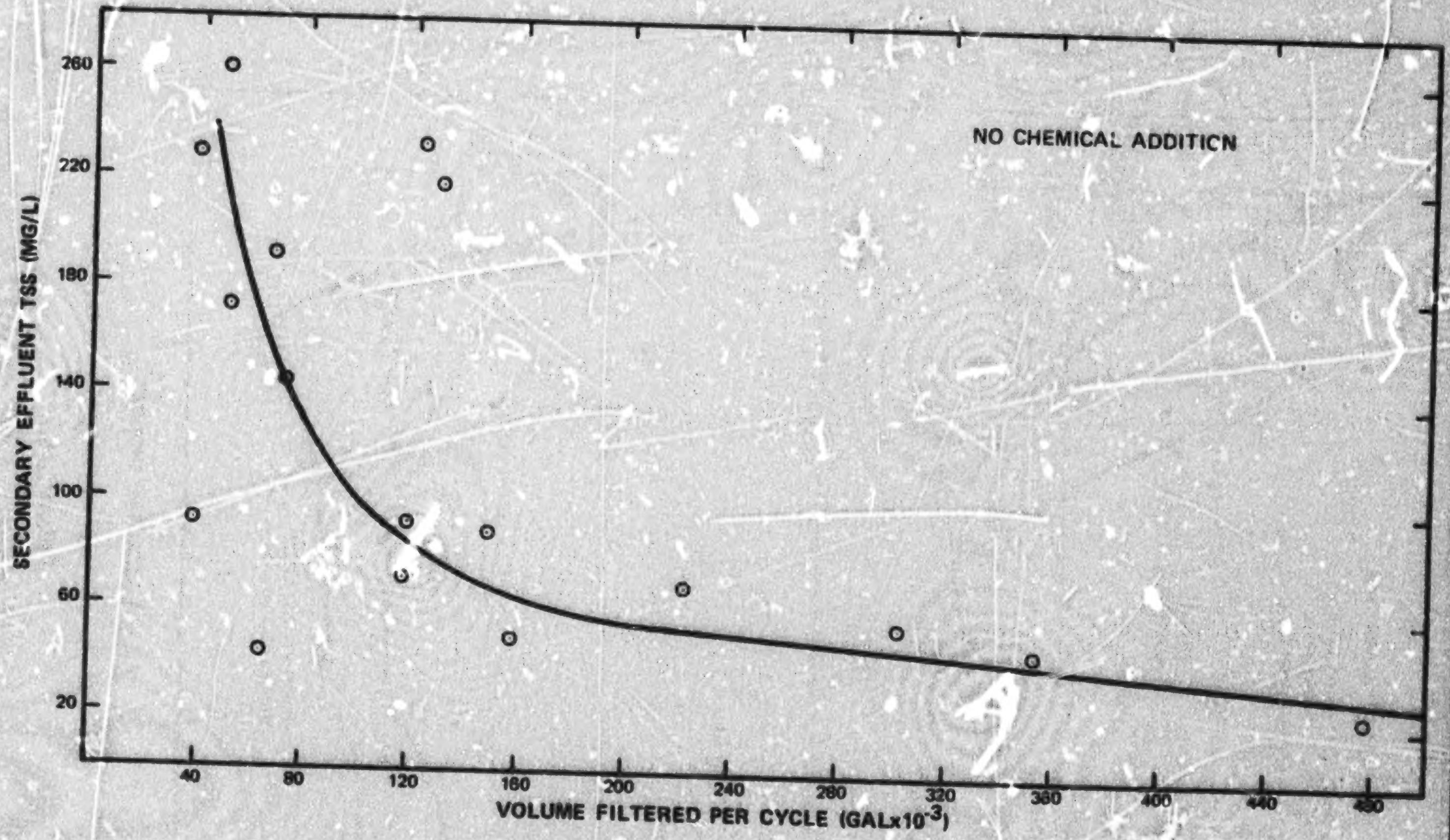


FIGURE 59

TERTIARY THROUGHPUT VOLUME VERSUS SECONDARY EFFLUENT TSS CONCENTRATION (WITHOUT CHEMICAL FEED)

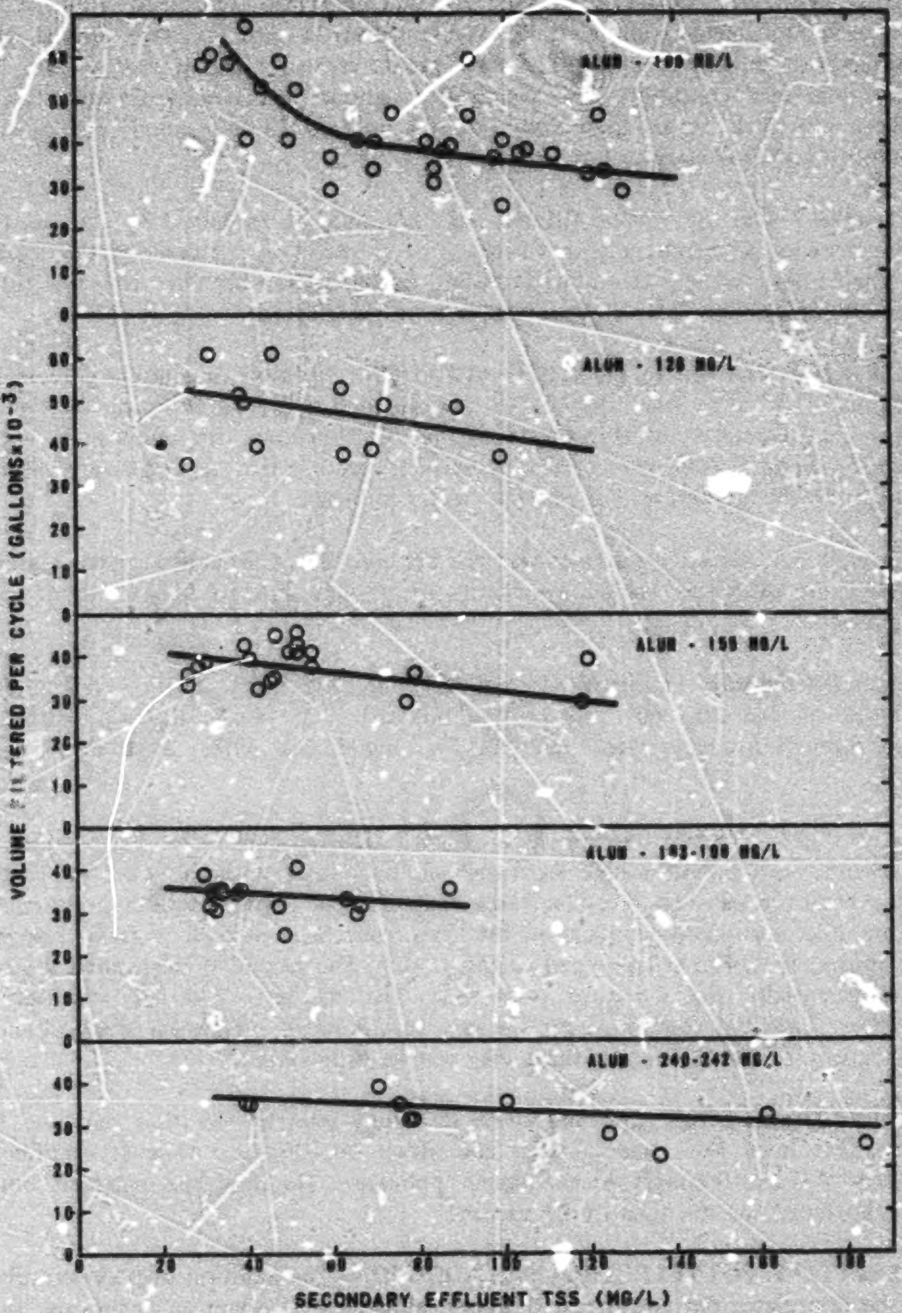


FIGURE 60

TERTIARY THROUGHPUT VOLUME VERSUS SECONDARY EFFLUENT TSS CONCENTRATION (WITH CHEMICAL FEED)

INFLUENCE OF ALUM DOSAGE ON CYCLE TIME–Increasing alum dosages also decreased the tertiary cycle time. Figure 61 is a plot of alum dosage versus the throughput volume per cycle. The data points represent the average volume filtered at each alum feed condition. The average throughput decreased nonlinearly from 50,000 gallons per cycle at an alum dosage of 53 mg/l to 32,000 gallons per cycle at alum dosages above 200 mg/l.

The reduction in the volume filtered per cycle may be explained by the fact that larger alum dosages produced greater quantities of chemical floc, increasing the solids loading on the tertiary settler/filter unit. Increased sedimentation capacity ahead of filter should tend to minimize the effect of alum dosage on filter cycle.

PREFILTER POLYELECTROLYTE ADDITION–Addition of anionic polyelectrolyte at dosages of 0.03 to 0.06 mg/l, after settling and prior to filtration, was effective in controlling turbidity breakthrough. Figure 62 illustrates the application of polymer prefilter. Three turbidity records were selected for identical alum and presettler polymer dosages, but differing prefilter polymer feed conditions.

In Figure 62a, prefilter polymer was not applied. Breakthrough occurred about midway through the filter run, with the turbidity increasing from 0.7 JTU to about 2.7 JTU at backwash.

The prefilter dosage in Figure 62b was 0.03 mg/l. Breakthrough occurred in the last 25 minutes of the run and was only about 0.1 JTU above the base turbidity level. Breakthrough was completely eliminated in Figure 62c with a prefilter dosage of 0.06 mg/l.

The turbidity control illustrated in Figure 62b is considered to be the optimum condition, with breakthrough occurring simultaneously with or just prior to backwash. While breakthrough was completely eliminated at the higher prefilter polymer dosage in Figure 62c, the higher application rate may have been at least partially responsible for the shorter filter run as compared to Figure 62b. The secondary suspended solids loadings were essentially the same for both runs. The alkalinity and pH condition differed, however. The feed conditions for the two runs were therefore not identical and a direct comparison of the filter cycle times may not be totally valid.

Prefilter dosages of 0.1 mg/l and above definitely reduced the filter run time. Figure 63 shows the head loss characteristics and effect on filter run time at prefilter polymer dosages of 0 and 0.1 mg/l. At the higher application, the head loss increased more rapidly and shortened the run time by 23 percent.

PRESETTLER POLYELECTROLYTE ADDITION–The addition of polyelectrolyte ahead of the flocculator did not appear to appreciably aid the formation or improve the settling characteristics of the chemical floc. Neither was there a noticeable increase in filter run time as a result of presettler polyelectrolyte addition.

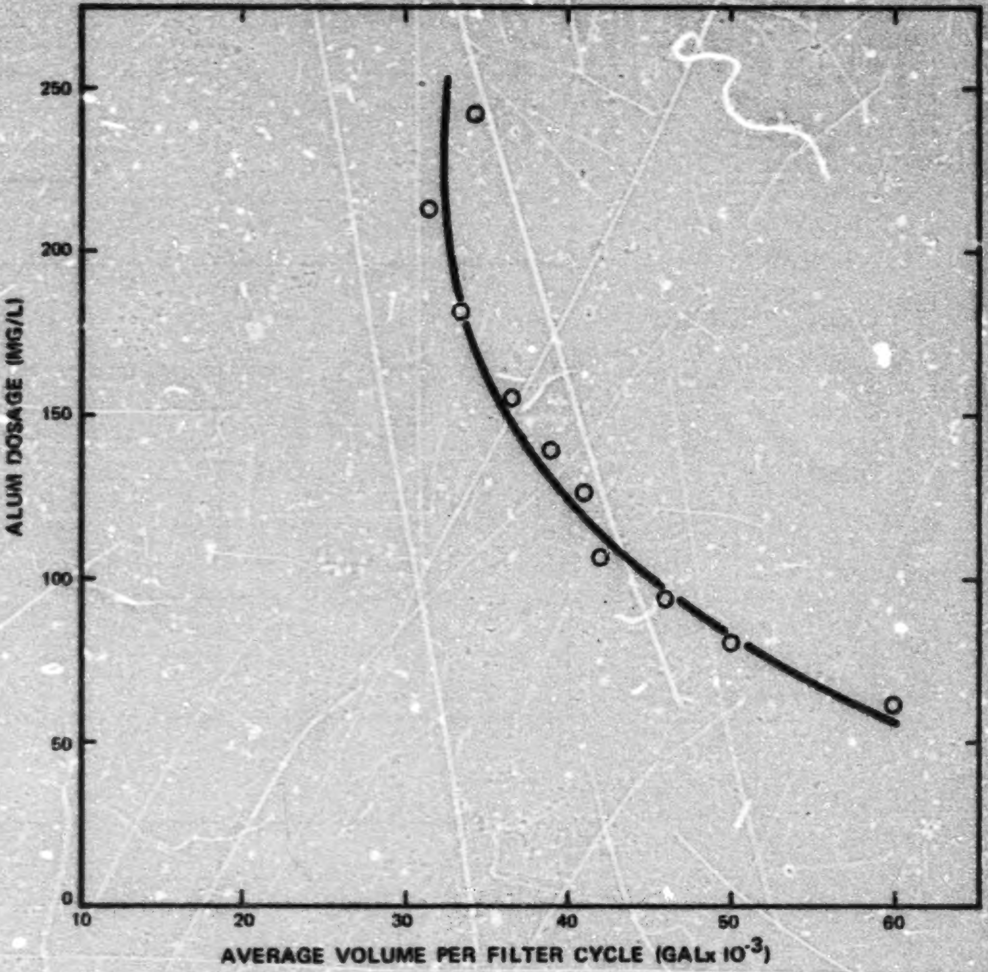


FIGURE 61

EFFECT OF ALUM DOSAGE ON FILTER CYCLE

DATE
21 APRIL 1971

CONDITIONS
FLOW RATE - 150 GPM
[illegible] - 126 MG/L
POLY. P/S - 1.4 MG/L
POLY. P/F - 0 MG/L
SEC. TSS - 60 MG/L
TER. pH - 5.0
TER. TSS - 0.4 MG/L

DATE
7 JULY 1971

CONDITIONS
FLOW RATE - 190 GPM
ALUM - 126 MG/L
POLY. P/S - 1.4 MG/L
POLY. P/F - 0.03 MG/L
SEC. TSS - 46 MG/L
TER. pH - 5.4
TER. TSS - 1.0 MG/L

DATE
2 JUNE 1971

CONDITIONS
FLOW RATE - 190 GPM
ALUM - 126 MG/L
POLY. P/S - 1.4 MG/L
POLY. P/F - 0.06 MG/L
SEC. TSS - 42 MG/L
TER. pH - 6.25
TER. TSS - <1 MG/L

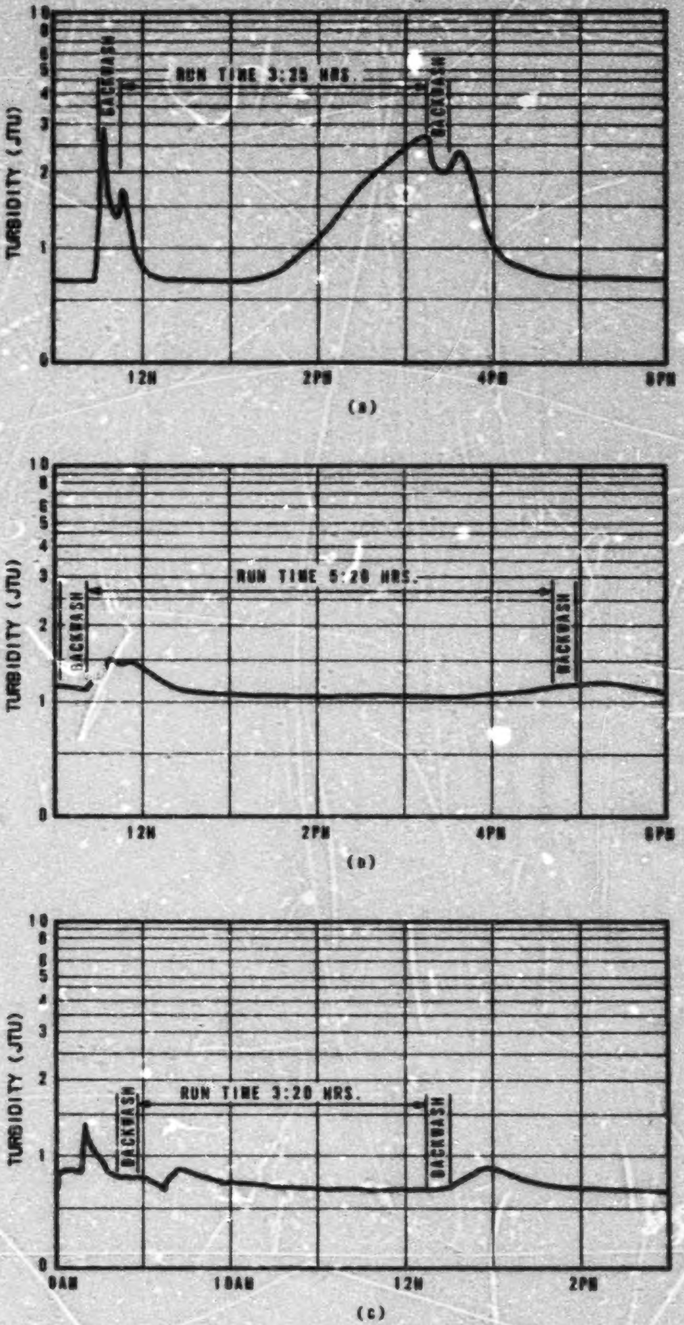


FIGURE 62

EFFECT OF PREFILTER POLYELECTROLYTE ADDITION
ON FILTER BREAKTHROUGH

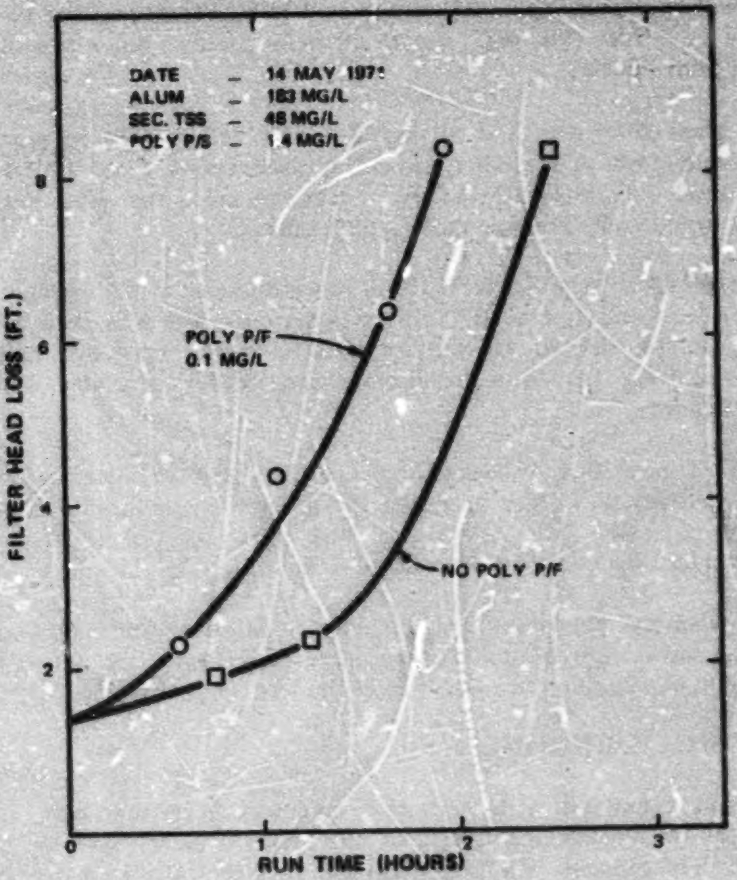


FIGURE 63

EFFECT OF PRE FILTER POLYELECTROLYTE ADDITION ON FILTER RUN TIME

Figure 64 is a plot of the secondary effluent TSS versus the pounds of secondary suspended solids removed per filter run for three alum feed conditions and three levels of presettler polymer addition - None, 0.5 to 1.0 mg/l, and 1.4 to 2.0 mg/l.

The data used in Figure 64 were limited to a prefilter polymer dosage of 0 - 0.06 mg/l to minimize filter "binding," due to excessive prefilter polymer addition and a reduction of the filter run time.

At alum dosages of 126 mg/l, and 155 to 160 mg/l, there was a slight increase in the solids removal at presettler polymer concentrations of 1.4 to 2.0 mg/l over the addition of alum only. From these data, it was concluded that presettler polymer additions did not substantially improve tertiary performance.

The presettler polymer feed stream was introduced at the constant head box between the tertiary influent pump and the flocculator tank. The alum feed stream was introduced on the suction side of the tertiary influent pump. Sufficient time may not have been provided for the alum to disperse and become hydrated before the polymer was introduced.

The chemical floc formed in the tertiary system was much smaller and appeared to be less cohesive than that produced in jar tests at the same dosages of alum and anionic polymer. From this observation, the point at which the presettler polymer was introduced may not have been the optimum location.

In summary, the results of polymer addition to aid the flocculation process were inconclusive. Further study is needed to better assess the potential benefits to be gained by polymer addition in this phase of the tertiary process.

PHOSPHATE REMOVAL

Phosphate removal in the tertiary system was observed to be a function of at least three variables: alum dosage, pH, and the initial concentration of orthophosphate in the secondary effluent. The percentage removals generally increased at higher alum dosages. The removal efficiency also improved with decreasing tertiary effluent pH. These observations were confirmed by jar tests conducted periodically during the project.

An attempt was made to relate all three parameters - pH, initial (secondary effluent) orthophosphate concentration and alum dosage - as shown on Figure 65. Log-log plots were made of the observed removal efficiency (tertiary ortho P/secondary ortho P) versus the molar ratio of Al (III) to secondary orthophosphate ion concentration for four pH ranges: 5.35 to 5.5, 5.8 to 6.35, 6.5 to 6.65 and 6.7 to 7.0. Straight lines were placed through the centroid of the data points and are replotted on Figure 66.

Although the data are somewhat scattered, the log-log relationship of removal efficiency to [Al(III)]/[Ortho P initial] appeared to hold up to a molar ratio of about 2 to 3. Above this range, the data varied widely.

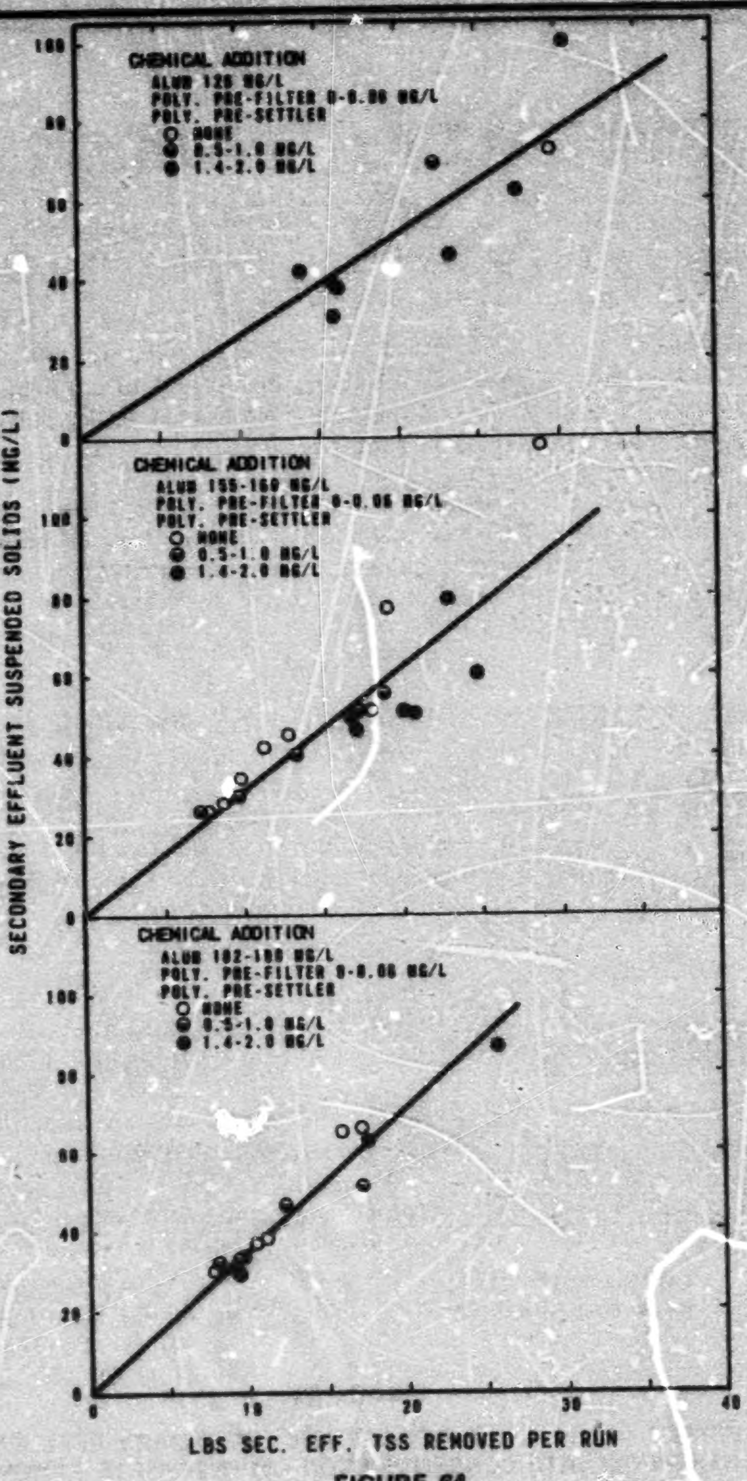


FIGURE 64

EFFECT OF PRESETTLER POLYELECTROLYTE ADDITION ON TERTIARY SOLIDS REMOVAL

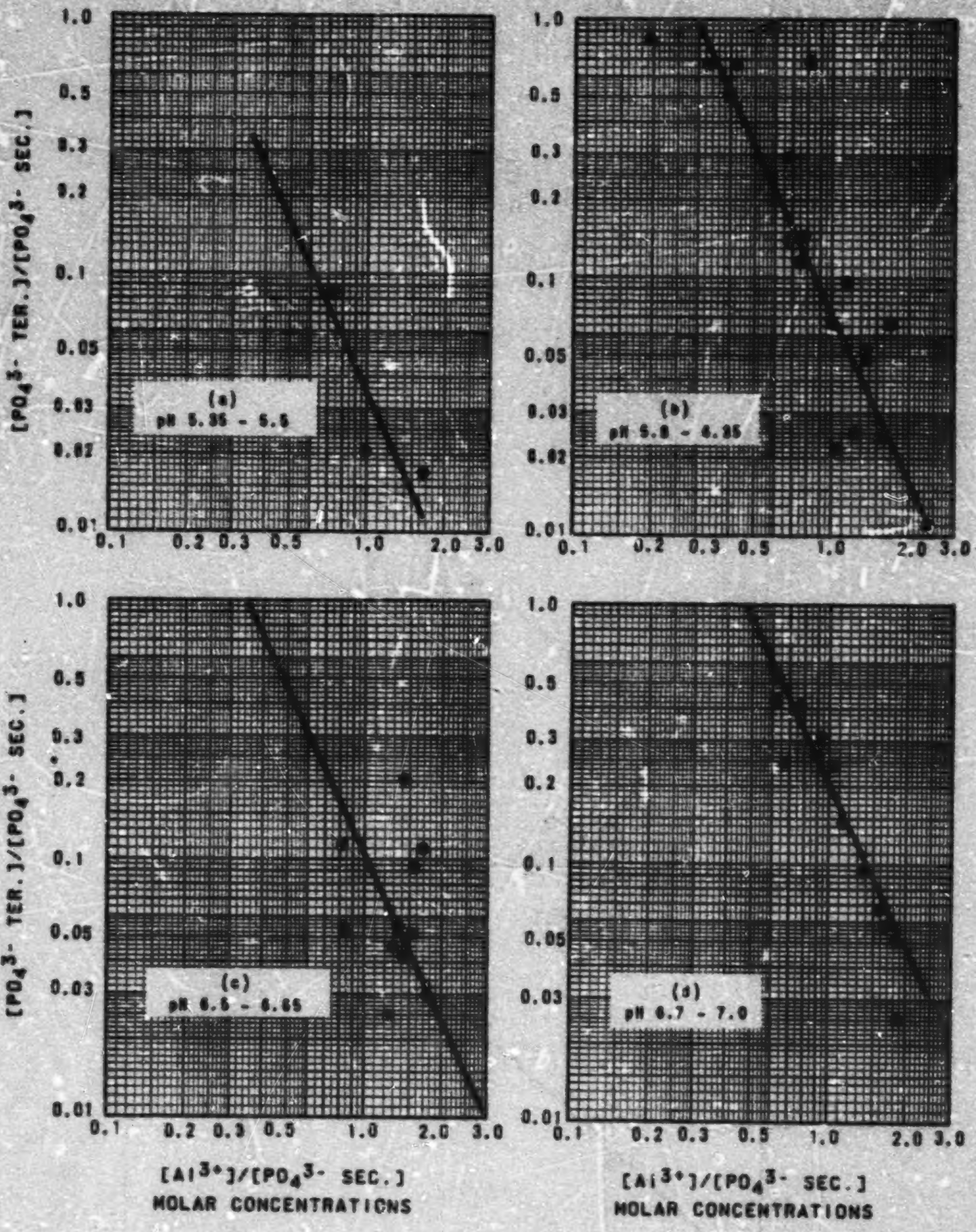


FIGURE 65

EFFECT OF pH, ALUM DOSAGE AND SECONDARY EFFLUENT ORTHOPHOSPHATE CONCENTRATION ON PHOSPHATE REMOVAL

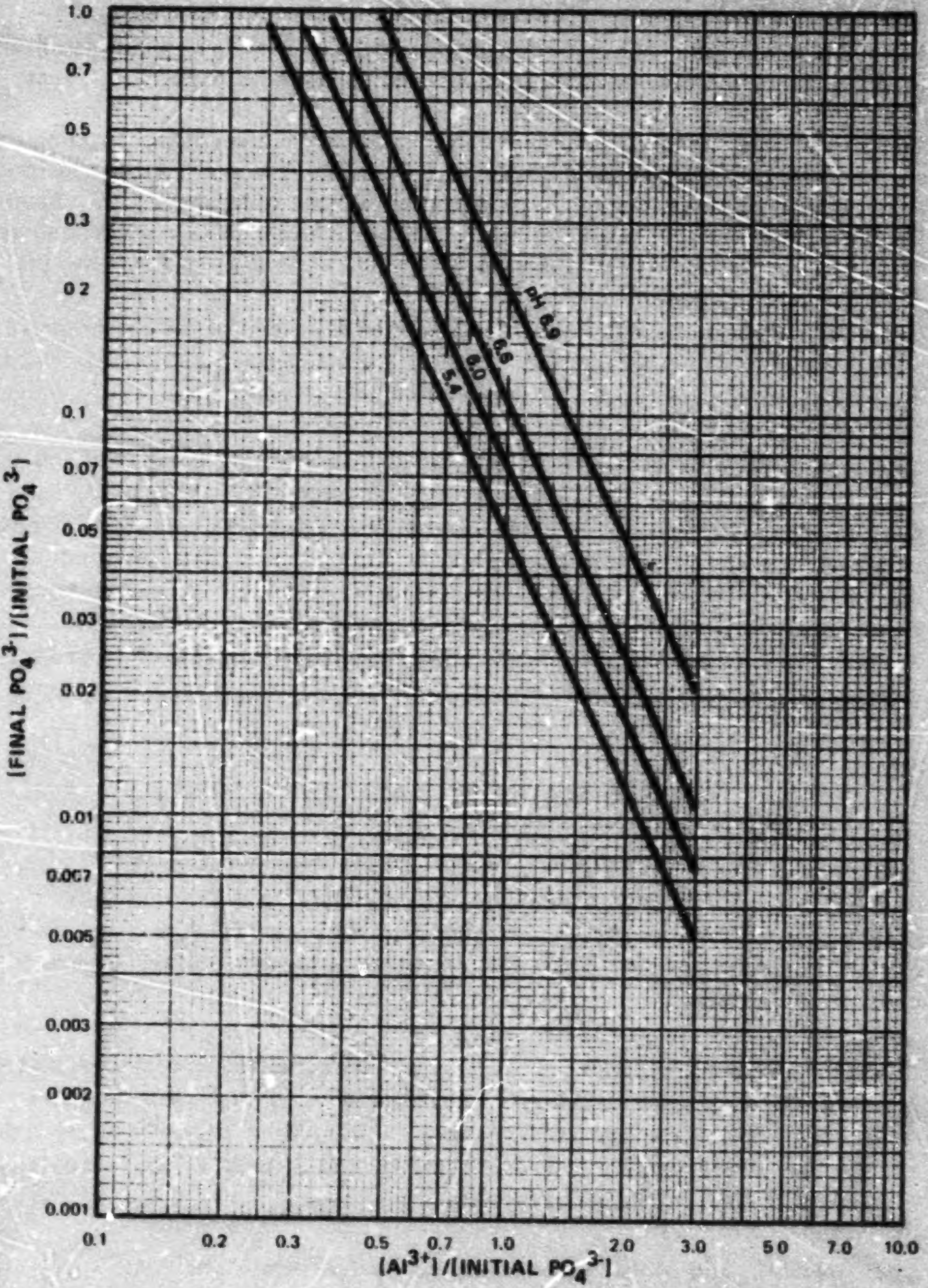


FIGURE 66

RELATIONSHIP OF ALUMINUM (III), INITIAL ORTHOPHOSPHATE CONCENTRATION AND pH ON ORTHOPHOSPHATE REMOVAL

From this observation, it was concluded that addition of alum at a molar ratio of Al(III) to initial orthophosphate in excess of 2 to 3 did not improve removal efficiency.

Lowering the pH from 6.7 - 7.0 to 5.3 - 5.5 reduced the alum dosage required to achieve a given orthophosphate removal by about 57 percent. Thus, a significant savings in alum addition can be realized by tertiary operation at the lower end of the effective alum flocculation pH zone. In addition, lower orthophosphate residuals can be achieved at the maximum effective [Al (III)]/Ortho P initial] molar ratio of 2 to 3 at lower pH.

Recht and Ghassemi [32] have also demonstrated the effect of pH on phosphate removal efficiency. They determined a molar ratio of [Al (III)]/[Ortho P initial] of 2 to be the upper limit of effective phosphate precipitation at a pH of 6.

Bicarbonate (HCO_3^-) alkalinity is also thought to have some influence on phosphate removal and may be responsible for some of the scatter in the data on Figure 65. The effect of this variable could not be quantitatively established, however. It is suspected that increasing concentrations of bicarbonate alkalinity at a given pH condition decrease orthophosphate removal efficiency. The orthophosphate and bicarbonate ions would be expected to compete with each other for adsorption sites on the alum floc, particularly as the pH increases. This is an interesting area of alum-phosphate chemistry, which deserves further study.

WASTE SOLIDS STORAGE AND DISPOSAL

The waste activated and chemical sludge storage lagoons were an effective means of retaining waste solids on the plant site. No objectionable odors were detected from the lagoons, even during the dewatering period at the end of the program. Unfortunately, ground water conditions limited the extent of solids dewatering, particularly in the chemical sludge lagoon.

Separate dewatering experiments using small drying beds demonstrated that both the chemical and waste activated sludge could be dried to a cinder-like humus material in 4 to 6 weeks, during the dry summer months. From these experiments, it appears that applying the stored solids to shallow drying beds at a depth of six to eight inches is an economical method for sludge dewatering.

The stabilized and dried waste activated sludge humus material is a suitable soil conditioner. Possible uses of this material include soil conditioning in the City's parks and application on selected agricultural croplands.

The dried chemical sludge from the storage lagoons may also be suitable as a soil conditioner. However, additional study is suggested to assess the possible adverse effects that may result from the aluminum present in the waste. In lieu of such tests, landfill is the recommended method for disposal.

DESIGN MODIFICATION AND EQUIPMENT RECOMMENDATIONS

GENERAL

Design modifications and items of additional equipment and instrumentation to improve the performance and operation of future secondary-tertiary treatment systems of the type demonstrated by this project are discussed in the following paragraphs. Several equipment design modifications made during the project have been discussed in previous sections. Additional sampling and testing equipment was not included in the original plant design, due to financial constraints. Considering the total capital investment of the treatment plant, the cost of the research and demonstration program, and the improvement in data quality and reliability that undoubtedly would have been attained, the minimal cost of this equipment would have been more than justified.

FLOW MEASUREMENT

The plant effluent was the only wastewater flow accurately measured and recorded. In plant designs where the aeration basin provides surge storage, as well as secondary treatment, it is recommended that the influent flow be continuously measured and recorded. The waste activated sludge should be metered and recorded to more accurately control wasting rates and sludge age. Where sludge lagoons are used for waste activated and chemical sludge storage, it is recommended provisions, such as a wier, be included to measure the supernatant return flow.

SAMPLING

For treatment systems receiving combined domestic and industrial wastewater with large fluctuations in both organic and hydraulic loadings, flow proportional automatic sampling of the influent, secondary effluent and tertiary effluent is recommended. It is also suggested that flow proportioned automatic samplers be installed to continuously monitor the significant industrial waste flows. Additional tertiary system sampling taps should be located in the flocculator effluent, between the tertiary tube settler and filter, and in the chemical sludge holding tank decant and concentrated sludge lines.

INSTRUMENTATION

Perhaps the most important control parameter for the tertiary process was pH. The extent of nitrification in the secondary system considerably influenced the secondary alkalinity and therefore, the tertiary effluent pH. It is recommended that the pH of the flocculator effluent be continuously measured and recorded. Where the size of the treatment facility can justify the cost, it is recommended that the aeration-surge basin D.O. and ammonium ion concentration also be measured and recorded continuously.

Both flow rate and filter head loss indicators were provided with the tertiary system, but neither parameter was recorded. A recording turbidimeter was borrowed during the demonstration program to monitor tertiary effluent turbidity. To effectively regulate chemical feed, assess tertiary system performance, and control plant effluent quality, it is recommended that tertiary filter head loss and turbidity be recorded, as well as indicated.

In larger installations where limitations are placed on phosphate discharge, it is suggested that automated monitoring and recording of the tertiary effluent orthophosphate concentration, with feedback control to the alum feed system, be considered.

INFLUENT PUMP STATION

Debris clogged the submersible influent pumps on an average of 2 to 3 times each month, requiring the units to be pulled and the impellers cleaned. A backhoe was brought in each time to lift the pumps out of the wet well for servicing. Where this type of pump is used in the influent pump station of small plants, it is recommended that a permanent hoist be installed to facilitate maintenance.

AERATION-SURGE BASIN

AERATION CAPACITY—In extended aeration plants where industrial wastes cause wide fluctuations in the influent organic loading, it is suggested that two speed, as opposed to single speed, mechanical surface aerators be considered. Also, that more units of lower horsepower be used than a lesser number of higher aeration capacity. This concept allows the operator considerable flexibility in adjusting the air supply to satisfy the incoming oxygen demand and can result in significant power savings. In larger installations feedback control of the aeration system from continuous D.O. monitoring may be justified.

GAS VENT SYSTEM—Where impermeable membranes, such as PVC, are used to seal the aeration-surge basin, it is recommended that a venting system to provide gas relief under the liner be included in the basin design.

SECONDARY TUBE CLARIFIER

The air sparging system, installed during the demonstration program, proved to be an effective means of cleaning the tube modules in the secondary clarifier. It is recommended that an air sparger be included in the design, if steeply inclined tubes are used for clarification of the secondary effluent. In addition it is recommended that provisions be made for thickening waste activated sludge in the secondary clarifier design.

TERTIARY SYSTEM

FLOCCULATOR–Considerable short circuiting occurred in the flocculator tank. The baffle modifications made midway through the project increased the mean detention time from about 50 to 75 percent of the theoretical value. In future tertiary systems of this type, it is recommended that careful consideration be given to preventing short circuiting in the design of the flocculator. For small tertiary systems of 50,000 to 300,000 gpd, it may be convenient to include the flocculator as a part of the tertiary tube settler/filter unit.

TERTIARY TUBE SETTLER–High velocity constrictions in the inlet system to the tertiary tube settler were suspected to cause a breakup of the fragile chemical floc. Modifications were made to the unit in an attempt to alleviate this problem. The effectiveness of these changes was difficult to assess, however. It is recommended that the inlet structure to the tertiary tube settler be redesigned to prevent turbulence from damaging the chemical floc, once formed in the flocculator.

The tertiary tube settler did not have adequate storage volume for the solids loadings encountered. The high suspended solids content of the secondary effluent undoubtedly contributed to this condition. It is recommended, however, that the effective surface overflow rate be reduced to about 35 gallons per day per square foot.

CHEMICAL SLUDGE HOLDING TANK SUPERNATANT–The supernatant from the chemical sludge holding tank was normally pumped to the comminutor basin (Figure 11) and combined with the influent flow to the aeration-surge basin. When the secondary system was in a highly nitrified condition, the alkalinity in the aeration-surge basin was not sufficient to buffer the acidity of the tertiary backwash water. As a result, the aeration-surge basin pH could be depressed to intolerable levels of 5 and below.

To alleviate this condition, the entire backwash flow was pumped to the chemical sludge storage lagoons. The increased lagoon supernatant return flow often contained high suspended solids concentrations caused by dense algae blooms. A significant portion of the algae appeared to pass through the aeration-surge basin placing an additional solids loading on the tertiary system.

As an alternative to either of these methods of disposal, it is recommended that the chemical sludge holding tank supernatant be returned, after settling, directly to the tertiary influent line. Analyses of the supernatant indicate a suspended solids content of 10 to 20 mg/l can be expected, following 3 to 4 hours of settling. The additional solids loading placed on the tertiary system would be minimal.

CHEMICAL FEED–Low secondary effluent alkalinity conditions, produced by nitrification in the aeration-surge basin, required the addition of soda ash to the tertiary influent to maintain pH control. It is recommended that the chemical feed system for secondary-tertiary treatment plants of this type also include soda ash mixing, storage and metering facilities.

Polyelectrolyte addition was not observed to improve alum flocculation. The point at which polyelectrolyte was added to the tertiary influent may not have allowed sufficient time for the alum coagulation reaction to reach completion. As a result, the polyelectrolyte may have reacted as a primary coagulant, rather than effecting interparticle bridging to aid flocculation. It is recommended that polyelectrolyte be introduced sufficiently downstream of the alum feed point to allow at least 2 minutes of reaction time. Soda ash should be introduced into the tertiary influent between the points of alum and polyelectrolyte addition.

LABORATORY EQUIPMENT

It is recommended that the following laboratory equipment supplement the list contained in Appendix C:

1. Dissolved Oxygen Analyzer
2. Muffle Furnace
3. COD Apparatus
4. Colorimetric Phosphate Test Kit
5. Specific ion probes for ammonium and nitrate ions or ammonium and nitrite-nitrate test kits
6. Jar Test Apparatus

OPERATIONAL RECOMMENDATIONS

GENERAL

The tertiary process was demonstrated to be an effective and economical process for removing phosphate and polishing the effluent from an extended aeration secondary treatment system. The most significant variable affecting tertiary performance was pH. Optimum phosphate removals were observed between pH 5.5 to 6.0. The minimum tertiary effluent turbidity and suspended solids concentrations occurred around pH 6.0.

The extent of nitrification in the aeration-surge basin had a significant effect on the operation of the tertiary system. Oxidation of ammonium to nitrate ion reduced the secondary effluent alkalinity allowing the acidity of the alum to depress the pH into a range more favorable to phosphate coagulation. Minimum tertiary effluent turbidity levels tended to occur when nitrate concentrations were between 5 to 20 mg/l.

Recommendations are offered for operation of the Tualatin secondary-tertiary system in the following paragraphs. In general, these suggestions are also applicable to future treatment systems of this type.

SECONDARY SYSTEM

MLSS–The MLSS should be held between 1,200 to 2,200 mg/l. Above and below these limits, the secondary tube clarifier performance can be expected to degenerate.

SLUDGE AGE–A sludge age of 5 to 6 days is sufficient to induce nitrification. Essentially, complete conversion of ammonium to nitrate ion will occur at sludge ages of 10 to 12 days, assuming adequate D.O. is available. Within the constraints imposed by the allowable range of MLSS, it is recommended that the sludge age be held to between 8 to 10 days. If the secondary alkalinity is not sufficient to offset the acidity of the tertiary alum dosage, the sludge age should be reduced by sludge wasting to limit the extent of nitrification.

D.O.–A D.O. level of 1.0 to 1.5 mg/l will support full nitrification. D.O. concentrations of 0.4 to 0.6 mg/l will inhibit nitrification and the SVI can also be expected to increase under these low D.O. conditions.

pH–Optimum performance of the secondary tube clarifier was observed between pH 6.9 to 7.2. The highest removals of phosphate in the secondary system tended to occur between pH 7.0 and 7.25. Insofar as practical, the aeration-surge basin pH should be held about pH 7.0 by controlling the extent of nitrification through regulation of the sludge age.

TERTIARY SYSTEM

ALUM ADDITION AND PHOSPHATE REMOVAL–Phosphate removal was determined to be a function of the secondary effluent orthophosphate concentration, alum dosage and tertiary effluent pH. These three variables are related as shown on Figures 65 and 66. To effect optimum phosphate removal from a given secondary effluent quality, it is recommended that the following procedure be followed:

1. Determine the total alkalinity of the secondary effluent.

2. Based on the experimentally determined acidity equivalent of 0.36 mg/l [equivalent $CaCO_3$] per mg/l alum [$Al_2(SO_4)_3 \times 14.3\ H_2O$], calculate an alum dosage which will leave 10 to 20 mg/l of total alkalinity in the tertiary effluent. This residual should be adequate to reduce the tertiary effluent pH between 5.5 to 6.3.

3. After the tertiary pH stabilizes, trim the alum dosage as required to maintain a tertiary pH of 5.9 to 6.1.

If a lower phosphate residual is desired and sufficient alkalinity is not available in the secondary effluent, add soda ash as necessary to increase the tertiary influent alkalinity to the level required.

POLYELECTROLYTE ADDITION–High molecular weight, medium charge anionic polyelectrolytes, were determined by jar tests to be the most effective flocculation aid to enhance alum floc formation. However, unless the reaction time between alum and presettler polyelectrolyte addition can be increased to 1-2 minutes, by relocating the points of chemical addition in the tertiary influent, it is recommended that presettler polymer addition be suspended.

Prefilter polyelectrolyte should continue to be added in an amount just adequate to prevent turbidity breakthrough (Figure 62). The dosage is expected to be in the range of 0.03 to 0.06 mg/l.

SAMPLING AND TESTING

It is recommended that the sampling and testing schedule used during the research and demonstration project continue to be followed. If the laboratory equipment recommended earlier in this section can be purchased, it is recommended that the following additional tests be performed:

ANALYSIS	SAMPLING LOCATION	FREQUENCY
O_2 Uptake Rate	Aeration-surge basin	Daily
MLVSS	Aeration-surge basin	3 per week
COD	Influent	3 per week
	Secondary Effluent	2 per week
	Tertiary Effluent	2 per week
Orthophosphate	Secondary Effluent	3 per week
	Tertiary Effluent	3 per week
Ammonium and Nitrate	Tertiary Effluent	3 per week

FUTURE RESEARCH AND DEMONSTRATION PROJECT RECOMMENDATIONS

Equipment and operational problems are inevitable in the startup of any research and demonstration program involving new equipment and technology. This project was certainly no exception. Difficulties encountered during startup and initial operation of the plant, required the duration of the program to be extended several months in order to fulfill the requirements of the EPA grant. Considerable additional expense to the grantee resulted from this extension.

It is recommended that for future research and demonstration programs of this type, a commissioning period of at least 3 to 6 months be allowed, as part of the overall project schedule, to check out equipment and establish the operating characteristics of the treatment system. This period is also necessary to familiarize the operator with new processes and equipment. Once stable operation of the system has been achieved, the scheduled testing and data acquisition phase of the project can proceed without repeated delays.

SECTION IX
FINANCIAL CONSIDERATIONS

The total capital cost of the treatment facility, including engineering, was $245,800. A detailed breakdown of the engineering and construction costs is included in Appendix D. The total capital cost of a conventional activated sludge treatment plant supplemented by chemical coagulation, sedimentation and filtration obtained from Smith's curves [38] and updated to January 1970 was $343,500. Comparison of these costs indicates the substantial capital cost savings of this approach to secondary-tertiary treatment.

OPERATION AND MAINTENANCE COSTS

The total operation and maintenance cost for the first year of service (August 1970 to July 1971) obtained from the City's accountant was $23,100. The cost includes allocation of a portion of the salaries of members of the City's public works and accounting staff, as well as the treatment plant operator.

Operation and maintenance costs for the plant, considering only the treatment plant operator's salary and a minimal allowance for accounting personnel to process plant records, was $19,400. This represents the actual annual cost to operate and maintain the treatment facilities. A detailed breakdown of the operational costs is included in Appendix D.

Operation of the tertiary portion of the plant was estimated to be $6,800, or 35 percent of the total operation and maintenance costs.

RESEARCH AND DEMONSTRATION COSTS

The cost associated with the research and demonstration program, including additional sampling and testing, supervision and data analysis was $31,800. This does not include the cost of the nitrogen and phosphorus analyses performed by the EPA Pacific Northwest Environmental Research Laboratory, or preparation of the final report.

TOTAL ANNUAL COSTS

The total annual costs of the combined secondary and tertiary treatment system for the first year of operation (August 1970-July 1971) are summarized in Table 14. The annual costs include:

1. Annual capital ammortization at a 5.5 percent interest rate over a 20 year period.
2. Annual operation and maintenance cost attributed to operation of the plant only during the first year of operation.

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3. Equipment replacement and depreciation costs estimated to be 3 percent of the equipment capital cost.

TABLE 14
TOTAL ANNUAL COSTS

ITEM		ANNUAL COST
Capital Cost		$245,800
Amortized Capital Cost (5.5% - 20 years)		20,600
Annual Operation and Maintenance		
Salaries	$10,200	
Utilities	4,500	
Chemicals	3,300	
Laboratory Supplies	600	
Insurance	800	
		$ 19,400
Estimated Equipment Replacement and Depreciation		3,000
TOTAL ANNUAL COST		$ 43,000

TREATMENT COSTS

At the beginning of the project, both the daily flow and organic loading were considerably below the design capacity of the plant. The average daily flow through the plant never reached the design condition. The design organic loading, however, was exceeded for a substantial part of the demonstration program.

On the basis of total annual cost and the average flow and organic loadings of 0.10 mgd and 467 lbs BOD_5 per day experienced over the demonstration project, the cost of treatment was calculated to be $1.18 per 1,000 gallons treated, or $0.25 per pound of BOD_5 removed.

The tertiary system, however, was operated at near the design flow rate for most of the project and adequately demonstrated the ability to provide the design level of treatment, even under severe loading conditions. Considering the low flows and wide range of organic loadings applied to the plant during the demonstration program, a more realistic calculation of treatment costs, which can be applied to future plants of this type, is one based on the plant design conditions.

At the design flow and organic loading of 0.28 mgd and 630 lbs BOD_5 per day, the cost for treatment of the combined industrial and domestic wastes at Tualatin was calculated to be $0.421 per 1,000 gallons treated, or $0.187 per pound of BOD_5 removed. The cost of the secondary treatment was estimated to be $0.287 per 1,000 gallons and tertiary treatment to be $0.134 per 1,000 gallons.

The following treatment costs were developed from Smith's curves [38], adjusted to January 1971 cost levels, the mid-point of the demonstration program:

Conventional activated sludge secondary treatment - $0.306/1,000 gal

Coagulation, flocculation and filtration - $0.197/1,000 gal

Total Treatment Cost - $0.503/1,000 gal

Comparison of the treatment costs developed from the R&D project with Smith's curves demonstrates the economy of this approach to providing tertiary treatment of combined industrial and domestic wastes for small municipalities.

Capital and treatment costs for the type of tertiary treatment facilities demonstrated at Tualatin, Oregon have been developed for design capacities ranging from 0.1 to 3.0 mgd in Figure 67, adjusted to June 1972.

The cost data in Figure 67 assume the following:

1. Alum dosage - 100 mg/l @ $0.03/lb
2. Polymer dosage - 0.05 mg/l applied prefilter only @ $1.90/lb
3. Chlorine dosage - 5 mg/l @ $0.15/lb
4. Power @ $0.012/kWh
5. One full-time operator for plant sizes of 0.1 to 0.5 mgd graduating to two full-time operators for a 2.0 mgd facility at the salary level indicated in Table 14.
6. Other operating and maintenance costs contained in Table 14.
7. Waste activated sludge and chemical solids are stored in lagoons at the treatment plant site. No costs were included for cleaning and disposal of the solids. These costs will vary depending on location and methods selected for ultimate disposal.

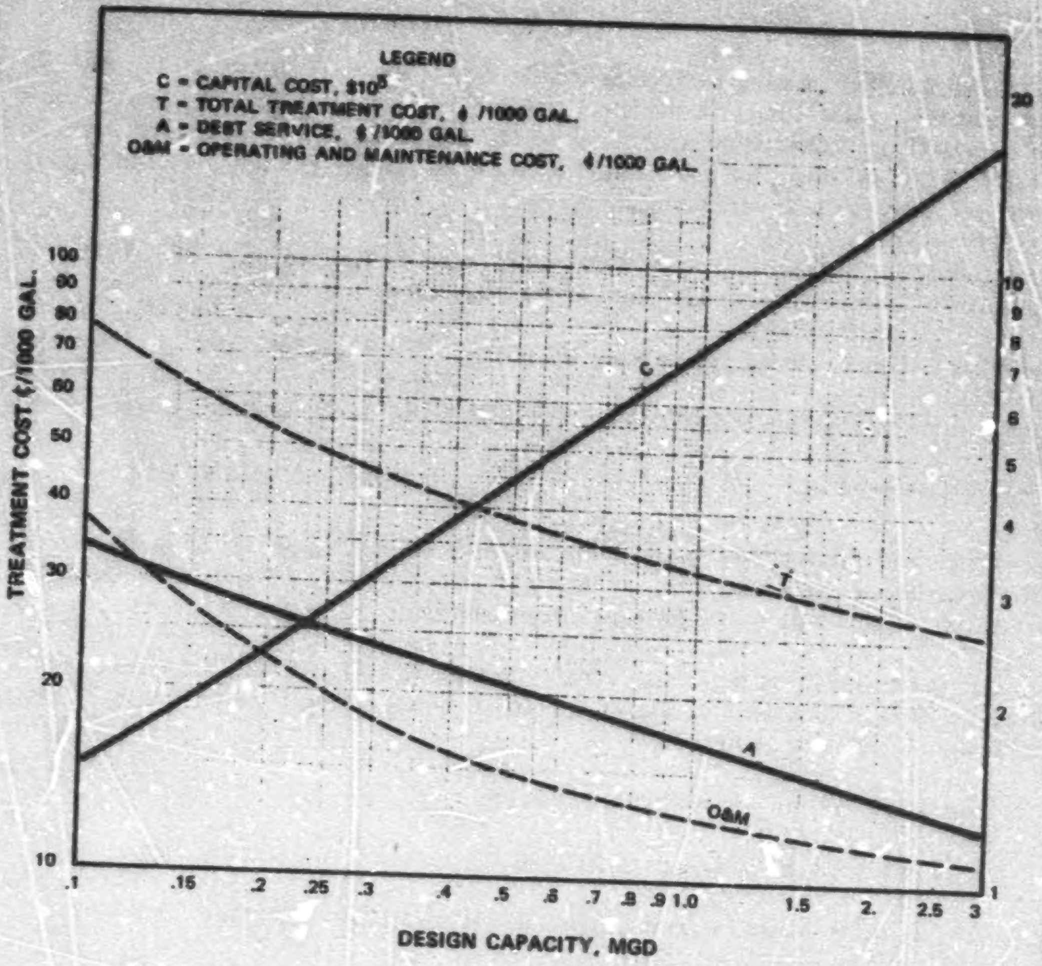

FIGURE 67

CAPITAL COST, OPERATING AND MAINTENANCE COST AND DEBT SERVICE VERSUS DESIGN CAPACITY ADJUSTED TO JUNE 1972

SECTION X
ACKNOWLEDGMENTS

This project was supported by the Environmental Protection Agency Research and Development Grant No. 11060 DLF. Appreciation is expressed to the administration and staff of the City of Tualatin, Oregon; particularly to the treatment plant operator-research technicians, Mr. Jim West and Mr. Dan Hanthorn; the EPA Pacific Northwest Environmental Research Laboratory personnel, Neptune MicroFLOC, Inc. who generously donated time and financial support to the project; and Cornell, Howland, Hayes and Merryfield/Clair A. Hill and Associates staff members for their cooperation and assistance during this study.

SECTION XI
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SECTION XIII
ABBREVIATIONS

mg/l – Milligrams per liter

BOD – Biochemical oxygen demand

MLSS – Mixed liquor suspended solids

MLVSS – Mixed liquor volatile suspended solids

mgd – Million gallons per day

TSS – Total suspended solids

VSS – Volatile suspended solids

SVI – Sludge volume index

D.O. – Dissolved oxygen

O_2 – Oxygen

T_s – Sludge age

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SECTION XIV
APPENDIXES

APPENDIX

A	Design Factors
B	Photographs
C	Process and Laboratory Equipment
D	Costs

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APPENDIX A

DESIGN FACTORS

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DESIGN FACTORS

INFLUENT PUMPS	
Number	2
Type	Submersible raw sewage, centrifugal
Capacity	250 gpm
Total Head	16.5 feet
HEADWORKS	
Sewage Grinder	
Number	1
Size	15-inch
AERATION BASIN	
Number	1
Depth (minimum)	12 feet
Average Volume (including clarifier)	280,000 gallons
Detention Time[1]	24 hours
Design Organic Loadings[2]	0.16 lb BOD_5/lb MLVSS/day (16.8 lb BOD_5/1000 cubic feet/day
AERATION EQUIPMENT	
Number of Aerators	2
Type	Floating, mechanical surface
Size	15 hp

SECONDARY TUBE CLARIFIER

Number	1
Surface Area	120 square feet
Surface Overflow Rate[3]	1.96 gpm/sq ft (2,820 gal/day/sq ft)
Solids Loading Rate[3,4]	47 lb/day/sq ft
Total Detention Time[3]	30 minutes
Number of Tube Bundles	2
Tube Bundle Overall Dimensions	20 ft (L) x 6 ft (W) x 3.5 ft (H)
Number of Modules per Tube Bundle	4
Tube Module Dimensions	10 ft (L) x 3 ft (W) x 3.5 ft (H)
Individual Tube Dimensions	
Cross-sectional Area (square, 2-inch/side)	4 square inches
Length	48.5 inches
Angle of Inclination (from horizontal)	60 degrees

WASTE ACTIVATED SLUDGE PUMP

Number	1
Type	Self-priming, centrifugal
Capacity	200
Total Head	8.7 feet

TERTIARY INFLUENT PUMP

Number	1
Type	Self-priming, centrifugal

TERTIARY INFLUENT PUMP (continued)

Capacity	235 gpm
Total Head	17.0 feet

FLOCCULATION TANK

Diameter	8 feet
Height	10 feet
Volume	3,540 gallons
Detention Time	15 minutes

TERTIARY TUBE SETTLER

Number	1
End Area Hydraulic Loading[3]	1.49 gpm/ft^2 (2,152 gal/day/sq ft)
Surface Overflow Rate[3]	143 gal/day/sq ft
Detention Time[3]	27 minutes
Number of Tube Modules	2
Tube Module Dimensions	9.25 ft (L) x 3.15 ft (W) x 8.5 ft (H)
Individual Tube Dimensions	
Cross-sectional Area (hexagonal, one-inch/side)	2.6 square inches
Length	39 inches
Angle of Inclination (from horizontal)	7-1/2 degrees

MIXED MEDIA FILTER

Number	1
Surface Area	49.5 feet^2

MIXED MEDIA FILTER (continued)

Design Filtration Rate[3]	4.75 gpm/ft^2
Backwash Rate	16 gpm/ft^2
Filter Media Depth	30 inches

Media Types	Anthracite Coal	Silica Sand	Garnet Sand
% of Total Bed Volume	55	30	15
Specific Gravity	1.5	2.6	4.0
Particle Size Range	1.0-1.2 mm	0.45-0.55 mm	0.2-0.3 mm

High Density Support Gravel	
Specific Gravity	4.0
Size	2.0 mm
Depth	3 inches
Low Density Support Gravel	
Average Specific Gravity	2.5
Size	3/16 to 2 inches
Depth	18 inches
Total Bed Depth (including support material)	51 inches

TERTIARY EFFLUENT PUMP

Number	1
Type	Centrifugal
Capacity	235 gpm
Total Head	35 feet

BACKWASH PUMP

Number	1
Type	Centrifugal
Capacity	800 gpm
Total Head	40 feet

SURFACE WASH PUMP

Number	1
Type	Centrifugal
Capacity	70 gpm
Total Head	180 feet

BACKWASH STORAGE TANK

Diameter	12 feet
Height	12 feet
Volume	9,600 gallons

CHEMICAL SLUDGE HOLDING TANK

Diameter	11 feet
Depth	15 feet
Volume	10,000 gallons
Settling Time	0 to 5 hours

CHEMICAL SLUDGE DECANT PUMP

Number	1
Type	Self-priming, centrifugal
Capacity	70 gpm
Total Head	13 feet

CHEMICAL SLUDGE PUMP

Number	1
Type	Self-priming, centrifugal
Capacity	120 gpm
Total Head	25 feet

ALUM STORAGE TANK

Diameter	8 feet
Height	11.5 feet
Volume	4,000 gallons

POLYELECTROLYTE AND SODA ASH FACILITIES

Mix Tank	250 gallons
Storage Tank	250 gallons
Auxiliary Storage Drum	55 gallons

CHEMICAL METERING PUMPS

Number	4
Type	Variable stroke, diaphragm
Capacity	540 gpd @ 25 psi

CHEMICAL SLUDGE STORAGE PONDS

Number	2
Area, each	0.32 acres
Depth	4 feet
Volume,each	264,000 gallons
Total	528,000 gallons

CHEMICAL SLUDGE STORAGE PONDS (continued)

Supernatant Drainage	Return to Plant Influent Pump Station
Operation	Decant Supernatant to Sludge Blanket and allow Sludge to Air Dry

ACTIVATED SLUDGE STORAGE PONDS

Number	2
Area, each	0.10 acre
Depth (maximum water level)	3 feet
Volume, each	42,000 gallons
Total	84,000 gallons
Supernatant Drainage	Return to Plant Influent Pump Station
Operation	Decant Supernatant to Sludge Blanket and allow Sludge to Air Dry

CHLORINATION

Type	V-notch chlorinator
Control	Tertiary Flow or Manual

CHLORINE CONTACT

Detention Time[1]	
Backwash Storage Tank	49 minutes
Outfall	16 minutes
Total	65 minutes

FLOW MEASUREMENT

Plant Flow	Prop. meter, plant effluent
Tertiary Flow Rate	Orifice meter, tertiary effluent
Waste Activated Sludge	Time clock, constant pumping rate

[1] At average design flow of 280,000 gpd.

[2] At 630 lb BOD_5/day and 2,000 mg/l MLSS.

[3] At tertiary design flow of 235 gpm.

[4] At 2,000 mg/l MLSS.

APPENDIX B

PHOTOGRAPHS

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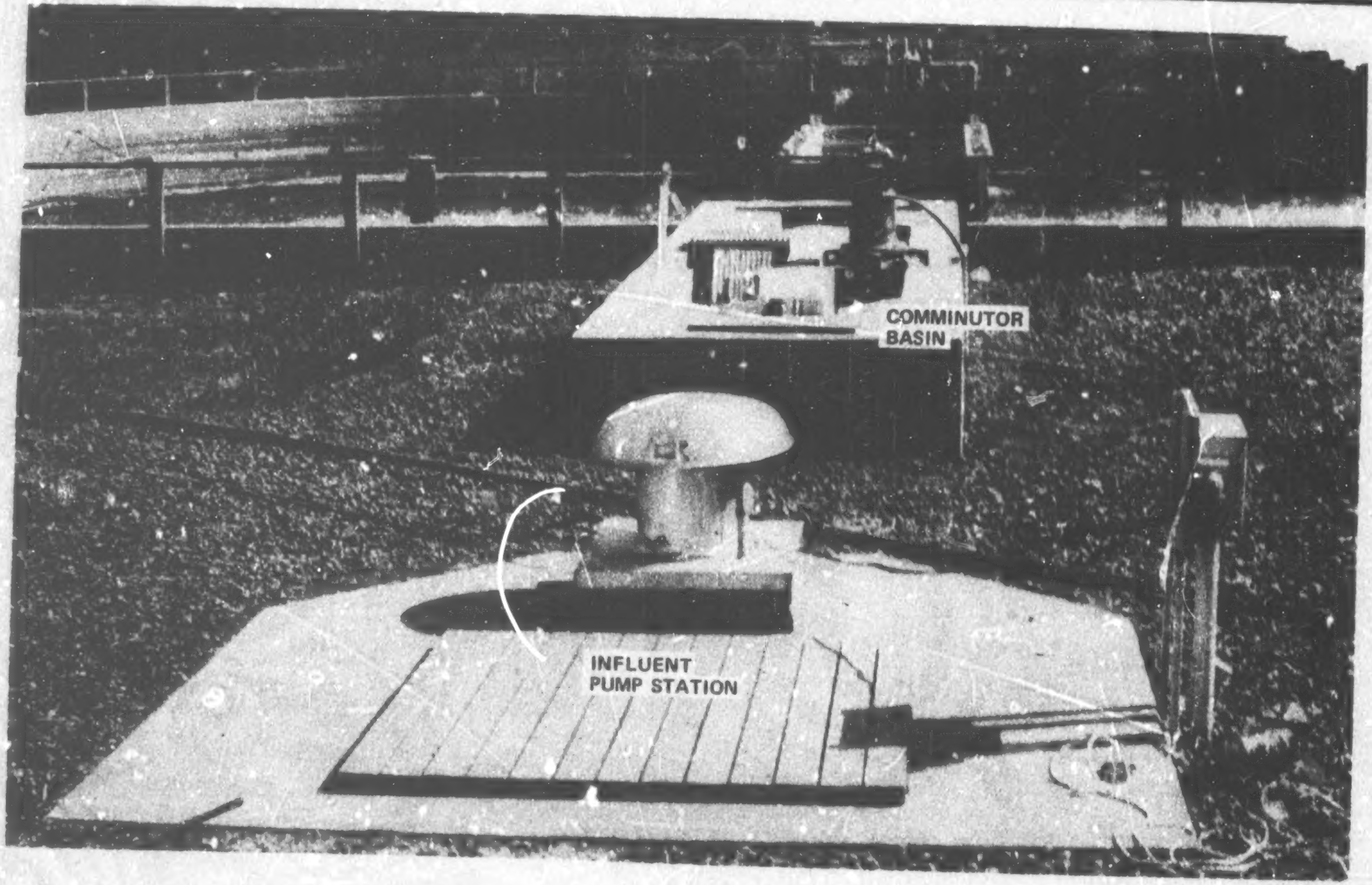
COMMINUTOR BASIN
INFLUENT PUMP STATION

FIGURE B –1

INFLUENT PUMP STATION AND COMMINUTOR BASIN

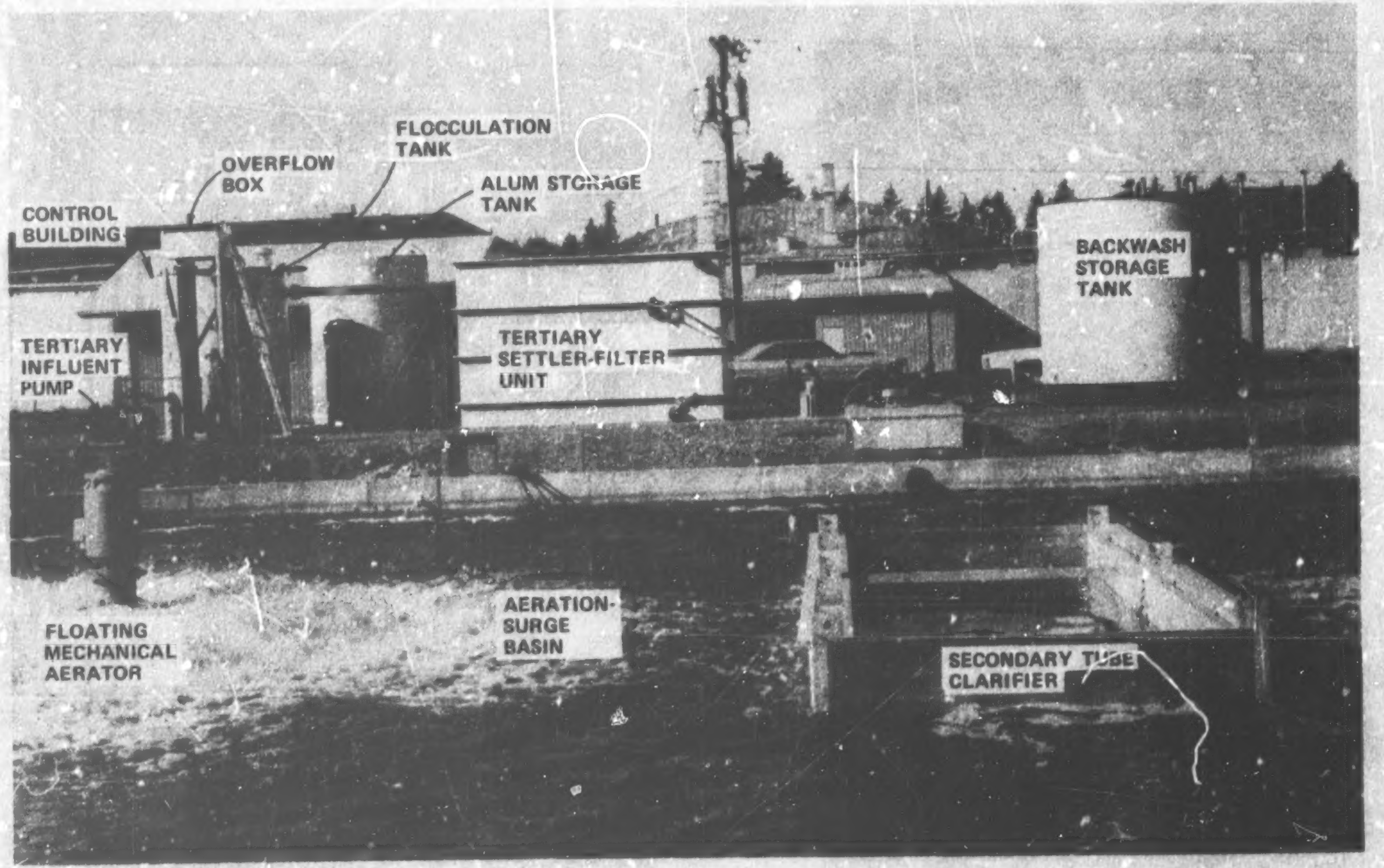
OVERFLOW BOX
FLOCCULATION TANK
ALUM STORAGE TANK
CONTROL BUILDING
BACKWASH STORAGE TANK
TERTIARY INFLUENT PUMP
TERTIARY SETTLER-FILTER UNIT
FLOATING MECHANICAL AERATOR
AERATION-SURGE BASIN
SECONDARY TUBE CLARIFIER

FIGURE B–2

AERATION-SURGE BASIN
AND TERTIARY SYSTEM

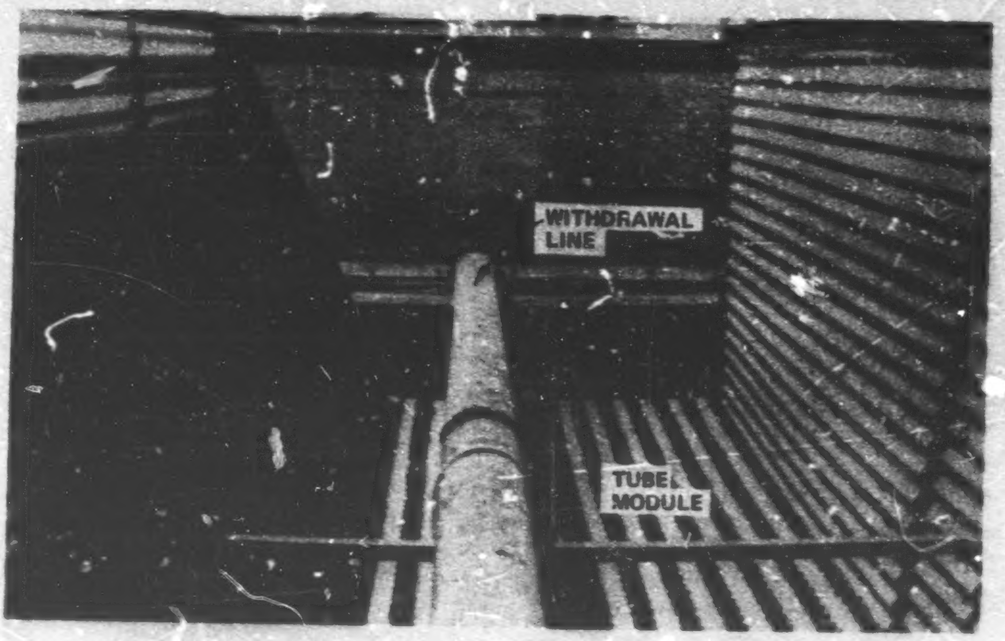
WITHDRAWAL LINE
TUBE MODULE

FIGURE B-3

SECONDARY TUBE CLARIFIER AND WITHDRAWAL LINE

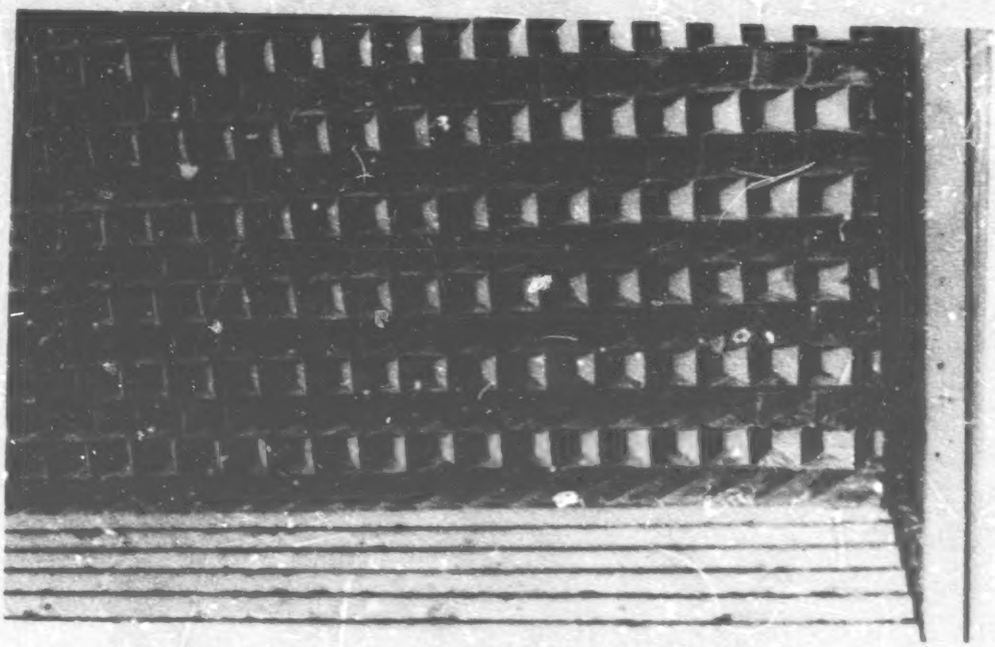

FIGURE B-4

SECONDARY CLARIFIER TUBE MODULE

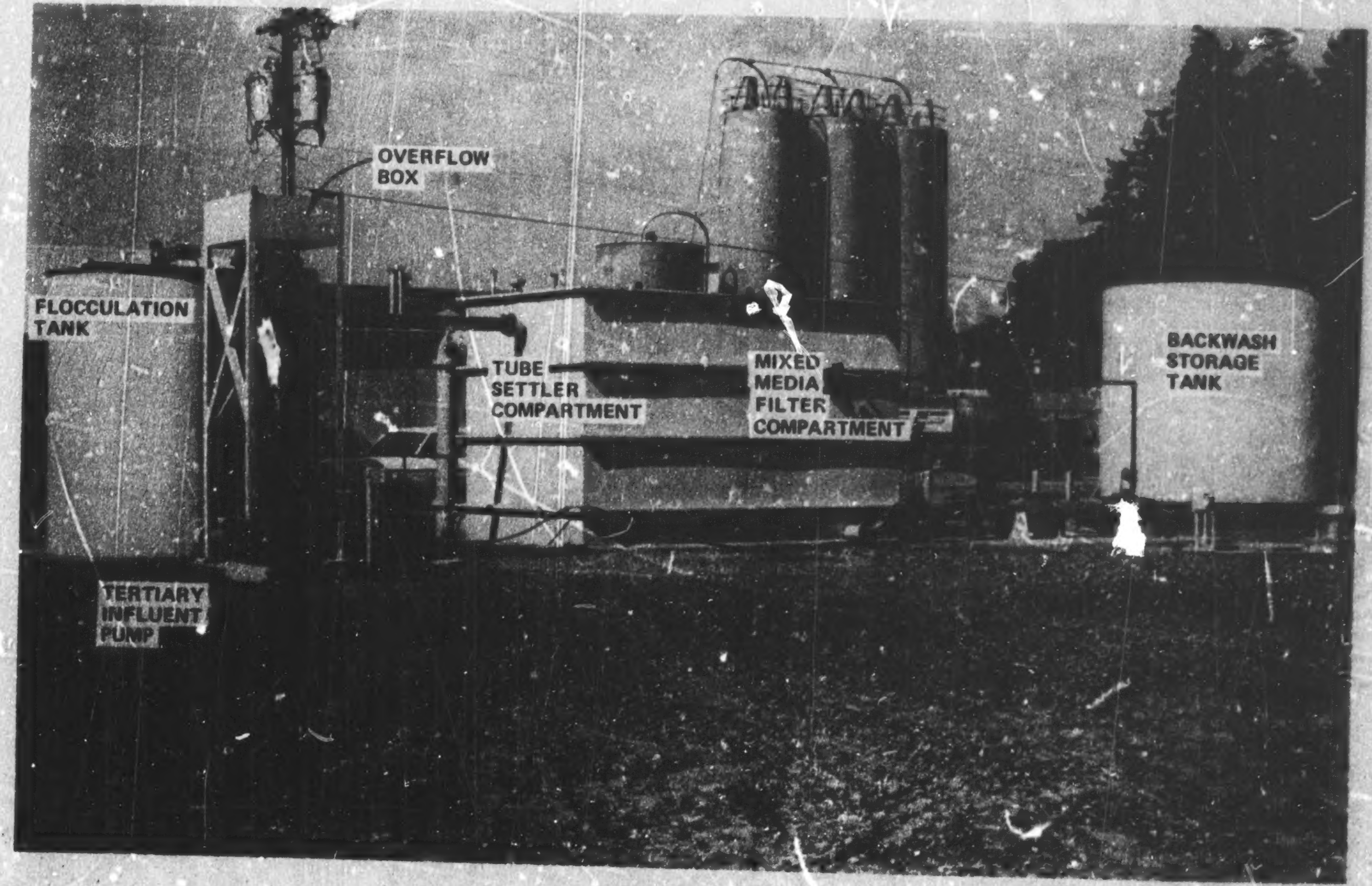
OVERFLOW BOX
FLOCCULATION TANK
TUBE SETTLER COMPARTMENT
MIXED MEDIA FILTER COMPARTMENT
BACKWASH STORAGE TANK
TERTIARY INFLUENT PUMP

FIGURE B–5

TERTIARY SYSTEM

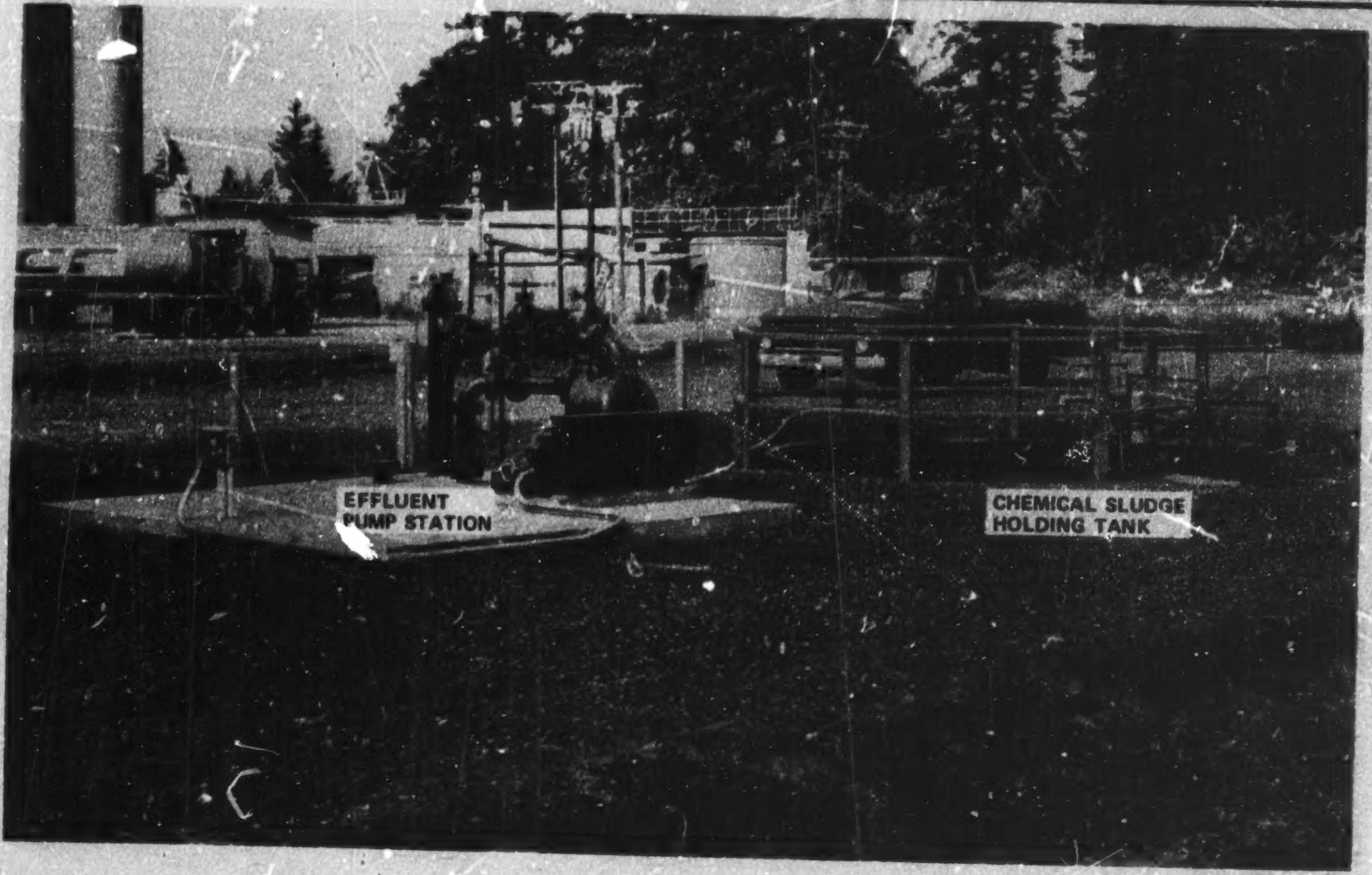
EFFLUENT
PUMP STATION
CHEMICAL SLUDGE
HOLDING TANK

FIGURE B-6

EFFLUENT PUMP STATION AND
CHEMICAL SLUDGE HOLDING TANK

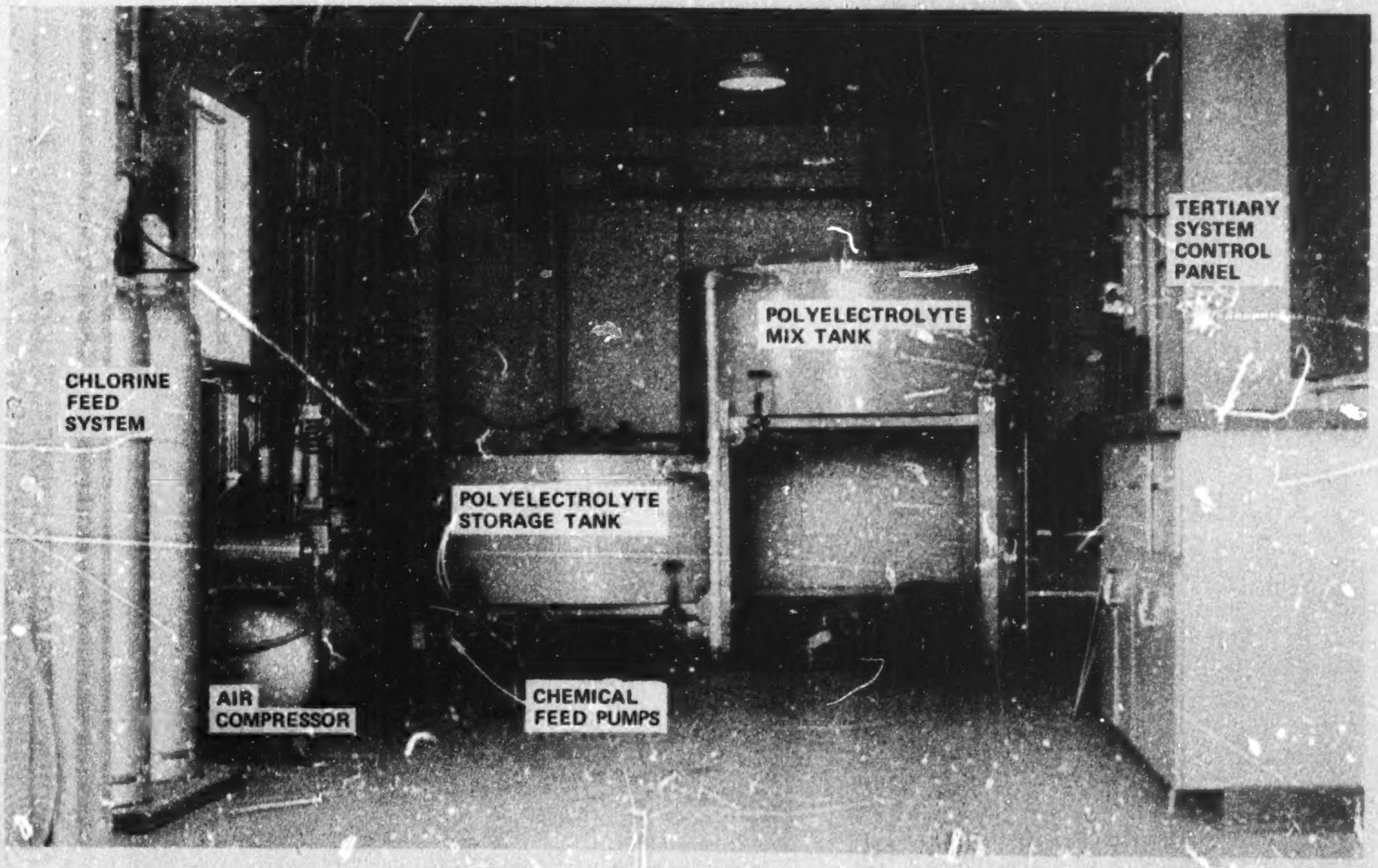
TERTIARY SYSTEM CONTROL PANEL
POLYELECTROLYTE MIX TANK
CHLORINE FEED SYSTEM
POLYELECTROLYTE STORAGE TANK
AIR COMPRESSOR
CHEMICAL FEED PUMPS

FIGURE B–7

CONTROL BUILDING

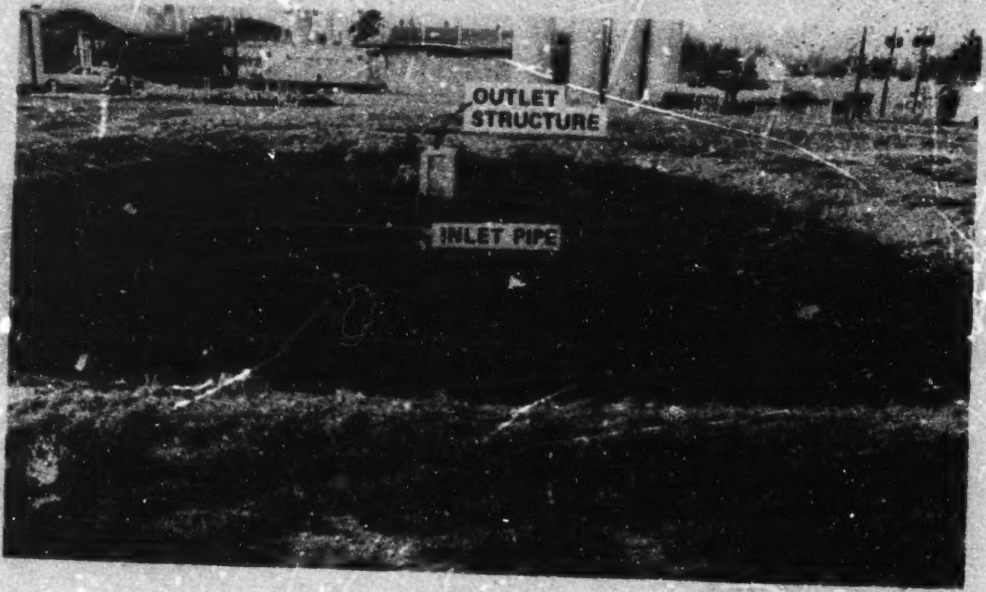
OUTLET STRUCTURE
INLET PIPE

FIGURE B–8

DEWATERED WASTE ACTIVATED SLUDGE
STORAGE LAGOON

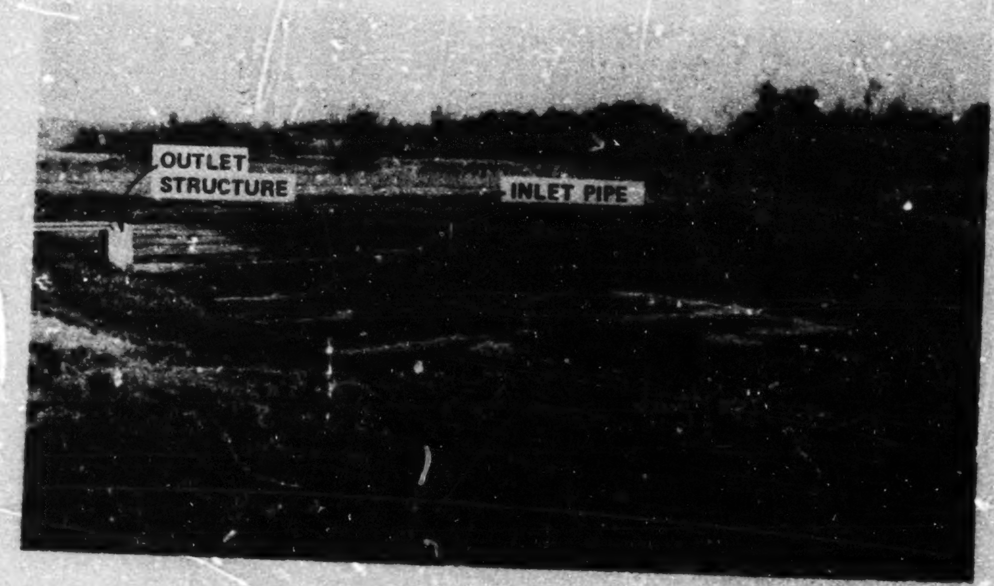
OUTLET STRUCTURE
INLET PIPE

FIGURE B–9

DEWATERED CHEMICAL SLUDGE
STORAGE LAGOON

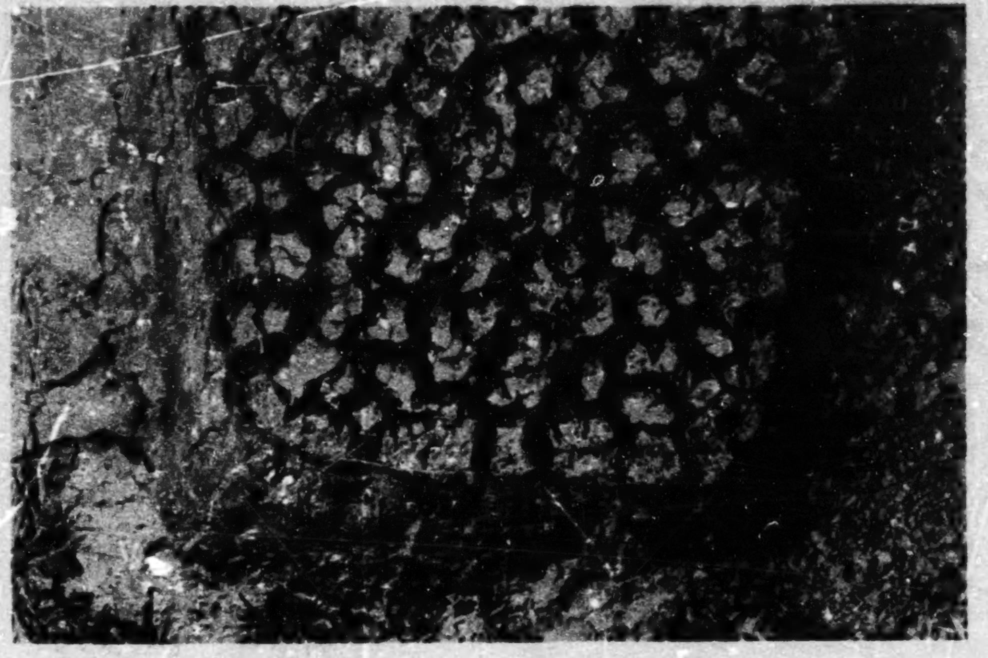

FIGURE B–10
DRIED CHEMICAL SLUDGE IN TEST BED

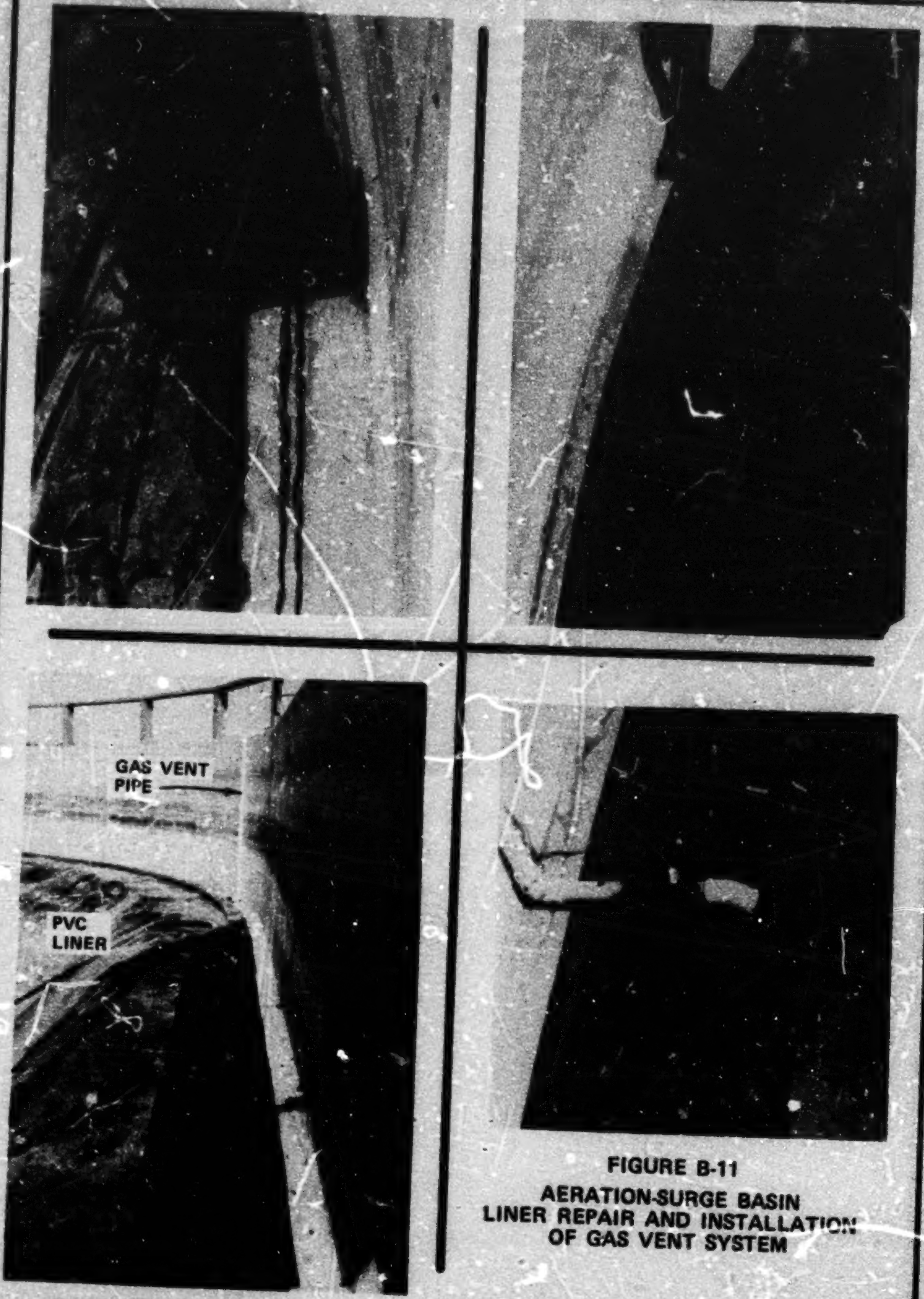
GAS VENT PIPE
PVC LINER

FIGURE B-11

AERATION-SURGE BASIN
LINER REPAIR AND INSTALLATION
OF GAS VENT SYSTEM

APPENDIX C

PROCESS AND LABORATORY EQUIPMENT

PROCESS EQUIPMENT

EQUIPMENT	MODEL	MANUFACTURER
Mechanical Surface Aerator	5K6237xJ290C15 Type K	Ashbrook
Activated Sludge Pump	T4A3-B	Gorman-Rupp Co.
Effluent Pump	T4A3-B	Gorman-Rupp Co.
Tertiary Influent Pump	13A2-B	Gorman-Rupp Co.
Chemical Decant Pump	12B2B	Gorman-Rupp Co.
Chemical Sludge Pump	12B2B	Gorman-Rupp Co.
Filter Effluent Pump	3655	Goulds
Surface Wash Pump	3655	Goulds
Backwash Pump	3655	Goulds
Raw Sewage Influent Pumps	CP3100	Flygt
Compressor	23ANLF4, Type 30	Ingersoll-Rand
Comminutor	12-4	Worthington
Secondary Tube Clarifier		Neptune MicroFLOC, Inc.
Tertiary Settler/Filter Unit		Neptune MicroFLOC, Inc.

LABORATORY EQUIPMENT

EQUIPMENT	TYPE	RANGE
Balance	Mettler HG Digital	0 - 160 g
Blendor	Waring	0 - 500 ml
Demineralizer	Barnstead Bantam	5 - 25 gal/hr
Drying Oven	Napco	30 - 200 degrees C
Filter Apparatus	Millipore, Pyrex	
Magnetic Stirrer	Magnestir	
pH Meter	Coleman, Metrion IV	0 - 14 pH
Refrigerator (2)	Westinghouse	
Refrigeration - incubator Conversion Unit	HACH Incutrol	0 - 35 degrees C
Vacuum Pump	Gast, Rotary Air Pressure and vacuum	1.3 cfm, 27 in Hg.

APPENDIX D

COSTS

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CAPITAL COST BREAKDOWN

ITEM		COST
Engineering		
Survey	$ 1,100	
Soils Investigation	550	
Engineering	18,900	
Services During Construction	4,200	
Resident Inspection	4,750	
		$ 29,500
Construction		210,100
Additional Construction and Equipment During Demonstration Program		4,900
Legal and Administrative		1,300
TOTAL CAPITAL COST		$245,800

CONSTRUCTION COST

ITEM	ESTIMATED COST
Move-In and Temporary Facilities	$ 550
Bond and Insurance	1,550
Influent Pump Station	9,500
Comminutor Basin	800
Aeration-Surge Basin	16,600
Aerators (2)	7,800
Secondary Tube Clarifier	9,000
Flow Measurement	1,050
Comminutor	4,550
Raw Sewage Pumps	3,800
Self Priming Pumps	5,000
Control Building	15,200
Yard Piping	7,500
Electrical	12,000
Sludge Ponds	9,000
Painting	3,050
Force Main	13,000
Cleanup and Finish Grading	3,150
Access Road	850
Fence	1,750

CONSTRUCTION COST (continued)

ITEM	ESTIMATED COST
Effluent Pump Station	8,300
Tertiary Treatment System	70,000
Laboratory Equipment and Reagents	3,000
Chlorination Facilities	1,450
Miscellaneous Concrete	1,650
TOTAL CONSTRUCTION COST	$210,100

FIRST YEAR
OPERATION AND MAINTENANCE COSTS
(August 1970 - July 1971)

ITEM		COST
Salaries		
Operator-Research Technician (Plant operation only - 3/4 salary)	$7,470	
Interim Operator(1)	680	
Operator Trainee(2)	1,690	
Public Works Superintendent(3)	2,930	
Accounting(3)	1,170	
Maintenance(3)	650	
		$14,590
Salary Overhead		770
Insurance		830
Utilities		
Power	3,320	
Water	800	
Telephone	360	
		$ 4,480
Chemicals		1,380
Laboratory Supplies		620
Miscellaneous Expenses		400
TOTAL FIRST YEAR O & M COSTS		$23,070

(See attached page for footnotes).

FOOTNOTES

1) Interim operator hired temporarily when operator-research technician was injured in an accident at the plant.

2) The operator-research technician left the employ of the City of Tualatin after the completion of the R & D Program. An operator-trainee was added to the staff five months prior to termination of the project.

3) The City of Tualatin allocated a portion of the salaries of the public works superintendent, accountant and public works maintenance staff as part of the operational cost of the wastewater treatment plant.

RESEARCH AND DEMONSTRATION COSTS
(April 1970 - August 1971)

ITEM	COST
Detailed Project Plan	$ 4,920
Operator-Technician (R&D Data Collection 1/4 Salary)	3,530
Consulting Engineering Services (Does not include preparation of final report)	21,770
Legal and Administrative	1,540
TOTAL RESEARCH AND DEVELOPMENT COST	$31,760

Note: Preparation of Final Report not included in R&D costs.

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